

-1-

PROCESS FOR SEPARATING AND RECOVERING ETHANE AND HEAVIER HYDROCARBONS FROM LNG

BACKGROUND OF THE INVENTION

[0001] The present invention example relates to the field of processing gasses in a liquid or fluid phase such as for LNG (Liquified Natural Gas) and NGL (Natural Gas Liquids) as known in the oil and gas industry and the recovery of C₂ and C₂+ (ethane +) components from the hydrocarbon fluid streams. More particularly, the present invention relates to the recovery of ethane and less volatile compounds from hydrocarbon fluid streams such as, from near atmospheric or more pressure, stored or transported cryogenic LNG liquid/fluid feed streams, with practical and economic design and operation of equipment to achieve this.

[0002] This invention relates to a process for separation of less volatile compounds, such as ethane and less volatile compounds from hydrocarbon mixed streams for example liquefied natural gas (LNG) or other such as petrochemical refinery streams. It is anticipated it could find utility in non-hydrocarbon related applications.

[0003] Background Art

[0004] Natural gas is being more often liquefied and transported in ocean going LNG tankers to LNG receiving terminals, worldwide. The LNG can then be re-vaporized and transported via pipelines carrying natural gas. The LNG can have other less volatile components besides the predominantly methane (methane usually makes up more than 50% of the LNG). It is usually necessary to remove various amounts of the less volatile components either to meet compositional specifications or Heating Value contractual terms, or in order to obtain greater value from the less volatile heavier compounds. This may be carried out at production, storage, loading terminals or receiving terminals. Storage of LNG presents the problem of uncontrollable "roll overs" caused by density inversions.

[0005] U.S. Patent No. 6,510,706 (Stone et al.) (January 28, 2003) discloses a process for removing hydrocarbons less volatile than methane from a pressurized liquid natural gas (PLNG). PLNG is heated in a heat exchanger, thereby vaporizing at least a portion of the PLNG. The partially vaporized PLNG is passed to a fractionation column. A liquid stream

-2-

enriched with hydrocarbons (C.sub.2+ or C.sub.3+) less volatile than methane is withdrawn from a lower portion of the fractionation column and a vapor stream lean in the hydrocarbons less volatile than methane is withdrawn from an upper portion of the fractionation column. The withdrawn vapor stream is passed to the heat exchanger to condense the vapor to produce PLNG lean in hydrocarbons less volatile than methane.

[0006] U.S. Patent No. 7,165,423 (Winningham) (January 23, 2007) discloses a process for the extraction and recovery of ethane and heavier hydrocarbons (C2+) from LNG. The process covered by this patent maximizes the utilization of the beneficial cryogenic thermal properties of the LNG to extract and recover C2+ from the LNG using a unique arrangement of heat exchange equipment, a cryogenic fractionation column and processing parameters that essentially eliminates (or greatly reduces) the need for gas compression equipment minimizing capital cost, fuel consumption and electrical power requirements. This invention may be used for one or more of the following purposes: to condition LNG so that send-out gas delivered from an LNG receiving and regasification terminal meets commercial natural gas quality specifications; to condition LNG to make Lean LNG that meets fuel quality specifications and standards required by LNG powered vehicles and other LNG fueled equipment; to condition LNG to make Lean LNG so that it can be used to make CNG meeting specifications and standards for commercial CNG fuel; to recover ethane, propane and/or other hydrocarbons heavier than methane from LNG for revenue enhancement, profit or other commercial reasons.

[0007] U.S. Patent No. 7,631,516 (Cuellar et al.) (December 15, 2009) discloses a process and apparatus for the recovery of ethane, ethylene, propane, propylene, and heavier hydrocarbons from a liquefied natural gas (LNG) stream is disclosed. The LNG feed stream is divided into two portions. The first portion is supplied to a fractionation column at an upper mid-column feed point. The second portion is directed in heat exchange relation with a warmer distillation stream rising from the fractionation stages of the column, whereby this portion of the LNG feed stream is partially vaporized and the distillation stream is totally condensed. The condensed distillation stream is divided into a "lean" LNG product stream and a reflux stream, whereupon the reflux stream is supplied to the column at a top column feed position. The partially vaporized portion of the LNG feed stream is separated into vapor and liquid streams which are thereafter supplied to the column at lower mid-column feed positions. The quantities and temperatures of the feeds to the column are effective to maintain

-3-

the column overhead temperature at a temperature whereby the major portion of the desired components is recovered in the bottom liquid product from the column.

[0008] U.S. Patent No. 7,216,507 (Cuellar et al.) (May 15, 2007) discloses a process and apparatus for the recovery of ethane, ethylene, propane, propylene, and heavier hydrocarbons from a liquefied natural gas (LNG) stream is disclosed. The LNG feed stream is divided into two portions. The first portion is supplied to a fractionation column at an upper mid-column feed point. The second portion is directed in heat exchange relation with a warmer distillation stream rising from the fractionation stages of the column, whereby this portion of the LNG feed stream is partially heated and the distillation stream is totally condensed. The condensed distillation stream is divided into a "lean" LNG product stream and a reflux stream, whereupon the reflux stream is supplied to the column at a top column feed position. The partially heated portion of the LNG feed stream is heated further to partially or totally vaporize it and thereafter supplied to the column at a lower mid-column feed position. The quantities and temperatures of the feeds to the column are effective to maintain the column overhead temperature at a temperature whereby the major portion of the desired components is recovered in the bottom liquid product from the column.

[0009] U.S. Patent No. 7,010,937 (Wilkinson et al.) (March 14, 2006) discloses a process for liquefying natural gas in conjunction with producing a liquid stream containing predominantly hydrocarbons heavier than methane is disclosed. In the process, the natural gas stream to be liquefied is partially cooled, expanded to an intermediate pressure, and supplied to a distillation column. The bottom product from this distillation column preferentially contains the majority of any hydrocarbons heavier than methane that would otherwise reduce the purity of the liquefied natural gas. The residual gas stream from the distillation column is compressed to a higher intermediate pressure, cooled under pressure to condense it, and then expanded to low pressure to form the liquefied natural gas stream.

[0010] U.S. Patent Publication No. 20080098770 (Ransbarger) (May 1, 2008) discloses a liquefied natural gas (LNG) facility employing an intermediate pressure distillation column for recovery of ethane and heavier components from the processed natural gas stream in a way that increases operational stability and minimizes capital and operating costs.

[0011] U.S. Patent Publication No. 20090221864 (Mak) (September 3, 2009) discloses that LNG is processed in contemplated plants and methods such that refrigeration

-4-

content of the LNG feed is used to provide reflux duty to the demethanizer and to further condense a vapor phase of the demethanizer overhead product. In such plants, the demethanizer provides a bottom product to a deethanizer, wherein a demethanizer side draw provides refrigeration to the deethanizer overhead product to thus form an ethane product and deethanizer reflux.

[0012] There are other processes to separate the heavier than methane hydrocarbons from LNG. However, the present innovation involves independent experimenting of all variations until arriving at a complete and a flexible viable process showing reduction of the design to practice by using the industry standard based design “HYSYS®” computer software Simulations tool (offered by Hypotech Company, Canada) which in addition supports the inventors’ “practical” equipment design, sizing and the operational information necessary for providing the required enablement’s for anyone skilled in the art or science.

[0013] Additionally, to address the issues presented by the prior art systems, the inventors believe that use of the present invention can be made to mitigate problems of the uncontrollable “roll overs” (caused by density inversions) in storage, through a process to separate and recover ethane from such LNG with ethane content as low as about 1% and lower. A part of the even leaner Lean LNG product as part of a liquid flash product of the Lean LNG product of this invention can be recycled to storage as part of a “roll over” control method. The separated heavier components have many uses such as for example, petrochemical feedstocks or liquid fuels.

BRIEF SUMMARY OF THE INVENTION

[0014] To address the forgoing desires, the present invention describes a process for separating and recovering ethane and heavier hydrocarbons from LNG. In one embodiment of the present invention the steps include providing an undivided feedstock stream containing Rich LNG wherein the Rich LNG is in liquid form from a storage tank or other source, the Rich LNG comprising C1 and C2+ hydrocarbons, the Rich LNG having an ambient storage temperature and pressure. The next step involves pressurizing the feedstock Rich LNG from storage pressure up to a desired pressure followed by pumping the feedstock Rich LNG into the cool side of a heat exchanger, the heat exchanger having a cool side and a hot side. The desired pressure is typically dictated by any downstream process steps involving the heat exchanger, and/or is dictated by critical pressure properties of the desired gas and liquid mixture in the column.

-5-

[0015] Then, the feedstock Rich LNG is heated within the heat exchanger while maintaining the feedstock Rich LNG below its bubble point to avoid vaporization while in the heat exchanger. In a preferred embodiment, the step of maintaining the feedstock Rich LNG below its bubble point to avoid vaporization while in the heat exchanger is achieved by regulating the pressure in the heat exchanger to maintain the Rich LNG in its liquid phase with no vaporization.

[0016] The undivided feedstock Rich LNG feed stream is directed from the heat exchanger to a processing column, the column comprising one or more stream entry ports along the height of the column to permit directing the stream into the column at one or more desired entry locations along the height of the column. The process then involves generating in the column a desired mixture comprising an overhead gas stream comprising lighter hydrocarbon products and a desired bottoms liquid stream comprising heavier hydrocarbon products. The overhead gas stream is directed from the column to the hot side of the heat exchanger. The next step involves cooling and condensing the overhead gas stream against the cold Rich LNG feedstock stream to form, in whole or in substantial part, a liquid comprising Lean LNG product stream, any remaining incidental uncondensed overhead gas stream remaining as a gas.

[0017] The condensed product stream is then directed from the hot side of the heat exchanger to a receiving vessel. The liquid Lean LNG product is pumped from the receiving vessel to a desired location. The bottoms liquid stream is directed from the column to one or more reboiler arrangements wherein heating of the bottoms liquid stream in the reboiler takes place. At least a portion of the heated bottoms stream are preferably returned to the column, the column being further outfitted with one or more heated bottoms stream entry ports along the height of the column to permit directing the heated bottoms stream into the column at one or more desired heated bottoms stream product entry locations along the height of the column.

[0018] The column bottoms stream is discharged directly from the column or from the reboiler and the bottoms stream is transferred to a desired location. Any gas in the receiving vessel is transferred to a desired location.

[0019] In another embodiment, this process may further comprise the steps of: directing the feedstock Rich LNG from the heat exchanger through a valve and into a degasser; directing the liquid stream from the degasser into the processing column, the

-6-

column being further outfitted with one or more degasser liquid stream entry ports along the height of the column to permit directing the degasser liquid stream into the column at one or more desired degasser liquid product entry locations along the height of the column; and directing any gas stream in the degasser to the column, the column being further outfitted with one or more degasser gas stream entry ports along the height of the column to permit directing the degasser gas stream into the column at one or more desired degasser gas product entry locations along the height of the column.

[0020] A portion of the column bottoms stream may be directed to the degasser to warm the feedstock and alter the composition of the total feed to the column.

[0021] The process may include the additional steps of recovering heat from the column bottoms stream.

[0022] In a preferred embodiment of the process, the NGL product comprises a desired high or low percentage of ethane.

[0023] In another embodiment, the Lean LNG stream may be directed to a storage facility or to further processing to vaporize the Lean LNG.

[0024] In one embodiment, at least some of the Lean LNG stream is directed to the column, the column being further outfitted with one or more Lean LNG stream entry ports along the height of the column to permit directing the Lean LNG stream into the column at one or more desired Lean LNG product entry locations along the height of the column.

[0025] The process may also comprise the additional steps of: directing at least some of the Lean LNG stream into one or more additional heat exchangers; heating the Lean LNG within the one or more heat exchangers while maintaining the Lean LNG below its bubble point to avoid vaporization while in the heat exchanger; directing the Lean LNG from the heat exchanger through a valve and into a degasser or other vessel; directing the liquid stream from the degasser or other vessel to a desired location; and directing any gas stream in the degasser or other vessel to a desired location. In one embodiment, the liquid stream may be directed from the degasser or other vessel into another heat exchanger arranged in series relationship and these additional steps be completed.

[0026] In one embodiment of the process, the Lean LNG stream is directed to a Rich LNG feedstock storage containing a level of Rich LNG feedstock, the storage further comprising one or more jet or sparger systems located along the height of the storage to

-7-

permit introduction of the Lean LNG stream into the storage either above and/or within the level of stored Rich LNG feedstock.

[0027] In another embodiment, the Lean LNG stream is directed to any stored source of LNG, wherein it is sparged into the stored source of LNG at a desired location.

[0028] In yet another embodiment, any gas phase in the receiving vessel is transferred to a compressor wherein the gas phase is compressed and then the compressed gas is directed to a desired location. The compressed gas may be directed into a heat exchanger wherein the compressed gas is condensed to form a full or partial condensate Lean LNG, the condensate then being directed to a desired location. The condensate Lean LNG stream may be directed to a storage facility. In one embodiment, at least some of the condensate Lean LNG stream is directed to the column, and introduced into the column via the one or more Lean LNG stream entry ports to permit directing the Lean LNG stream into the column at one or more desired locations along the height of the column. The heat exchanger may be cooled by an external refrigeration stream or by a second LNG stream.

[0029] In one embodiment, of the process, a second cold LNG stream is introduced directly into the degasser to mix with the feedstock Rich LNG. In another embodiment, the second cold LNG stream may be introduced directly into the column, the column being further outfitted with one or more LNG stream entry ports along the height of the column to permit directing the LNG stream into the column at one or more desired LNG stream product entry locations along the height of the column.

[0030] In one embodiment of the operation of the process, the step of cooling and condensing the overhead gas stream against the cold Rich LNG feedstock stream does not form any incidental gas.

[0031] In one embodiment of the process, the discharged bottoms stream comprises up to 99% of the C2 hydrocarbons in the Rich LNG feedstock and substantially all of the C3+ as a NGL, the NGL product additionally meeting without any further processing close to or substantially a Pipeline Quality Specification of = or < 0.5%v C1.

[0032] In another embodiment, the discharged bottoms stream comprises an NGL Product of substantially with TVP of up to = or < 400 psig, up to C1= or <0.5%v, up to 51%mol C2 or more fraction.

-8-

[0033] In yet another embodiment, the Rich LNG feedstock comprises between 1% mole C2 to that exceeding 40 to 50 mole % C2.

[0034] The process of the present disclosure may run in a high “ethane recovery” (90% or more) mode to recover up to 99% ethane and substantially 100% propane.

[0035] In one embodiment, the processing column employed comprises about 10 theoretical trays. The column has flexible configurations and preferably is configured and integrated in a multitude of operability and functional configurations selected from the group consisting of distillation columns, extractive distillation columns, reboiled absorption columns, absorption columns, lean oil absorber columns, fractionation columns, stripping columns, refluxed stripping columns, and reboiled stripping columns.

[0036] In one embodiment, substantially no tail gas (gas from condensed overhead stream) is formed even where there is as low as 1% C2 in the feedstock.

[0037] In operation of the process, NGL of Pipeline Quality specs is produced, even when the system is operating in deep high ethane extraction mode (90% plus).

BRIEF DESCRIPTION OF THE SEVERAL VIEWS OF THE DRAWINGS

[0038] FIG. 1A is a flow diagram of a HYSYS Simulation of a LNG processing plant in accordance with the present invention.

[0039] FIG. 1B is the detail area 1B shown in FIG. 1A.

[0040] FIG. 2A is another flow diagram of a HYSYS Simulation of a LNG processing plant in accordance with the present invention adding additional processing options to those described in connection with FIG. 1A.

[0041] FIG. 2B is enlarged view of detail area 2B shown in FIG. 2A.

DETAILED DESCRIPTION OF THE INVENTION

[0042] The present invention is generally concerned with the practical recovery or separation of less volatile components from a mix of other components as for example in this instance methane rich stream is separated from a stream of less volatile components than methane as may consist in LNG and such streams.

[0043] Also disclosed are designs for a novel but practical method and an arrangement, management and control of it to achieve such separation while at the same time

-9-

providing means and direction for achieving practical sizing and/or design and/or operation of the equipment used in performing the separations.

[0044] It further is designed and shown as for in this instance to take a “rich” in varying degrees, in lower volatile components than methane in LNG (termed “rich LNG”) stream, from its storage or transport conditions of but particularly liquid form, as far as possible maintain that phase or form while cross exchanging its cold energy condition to absorb heat from one of the product streams particularly the lean in the less volatile components LNG (termed “lean LNG”) stream obtained from equipment further downstream such as the vapor overhead stream from the processing column or by heating by other means of heating. This is a particular difference with the present innovation from other existing prior art offered or provided by others, in that the other prior art provides a design criteria or direction or instructions requiring a portion of the Rich LNG feed stream or splits of the Rich LNG stream be vaporized before feeding it to the processing column. As demonstrated for the present invention in the normal Ethane Recovery mode and provided result tables, there is no vaporization of the Rich LNG feed stream (stream 2) until the system is operated in an Ethane Rejection mode.

[0045] In other words, heat is exchanged in this heat exchanger (herein identified as LNG exchanger) while maintaining a liquid phase for the rich LNG stream and at the same time condensing the lean LNG vapor or mixed phase stream obtained from the processing column arrangement shown downstream from the LNG exchanger; and more particularly not vaporizing it in the LNG exchanger as one of the criteria for overall control, thereby resulting in a practical and reasonably sized heat exchanger. The type of heat exchanger can be any of the heat exchanger systems known in the art. The reference to a heat exchanger can include an individual heat exchanger or a multitude of individual heat exchangers. One manner in which to maintain the mixture in its liquid state is to maintain the mixture substantially or discernibly below bubble point. This can be achieved via the regulation of the pressure in the exchanger, e.g., by pressurizing the feedstock stream with a pump (or other motive power system) prior to entry into the exchanger and by maintaining sufficient back pressure in the system downstream of the exchanger (e.g., at the column or other location or pressure restrictor) to maintain the desired pressure in the exchanger. Maintaining the desired pressure in the exchanger permits the system to maintain liquid regimes for LNG coming in and LNG leaving the exchanger. This provides economy by conserving pressuring energy both in

-10-

terms of avoiding the use of unnecessary equipment and regarding the end PIPELINE gasifying/compressions.

[0046] A provision is provided to degas the liquid stream if necessary prior to feeding the warmed rich LNG to the processing column as a safety feature or a particular operational feature, prior to feeding the liquid stream at a top or high position in the processing column.

[0047] In the processing column the more volatile part of LNG is stripped from the less volatile part.

[0048] Further, the present invention avoids the usual or particular prior art instructions or necessity of pre-vaporizing in part or such and/or of splitting the feed stream prior to or after the LNG heat exchanger and/or having to have any part of or the whole feed stream pre-vaporized prior to feeding the processing column.

[0049] Further, the invention provides method/process/system/operations/means for practical and/or simplicity of design with resultant economy of equipment and/or utilities and/or operations.

[0050] This invention makes the design practicable and/or economic whereas some prior art designs or systems or methods or process(es) require additional utilities or equipment or as demonstrated have equipment design tending towards infinite or indeterminate heat exchange surface areas required of LNG exchanger design or have impractical temperature crosses in the LNG exchanger.

[0051] Further, the invention provides method/process/system/operations/means for design and operation in particularly eliminating compression/recompression of the Lean LNG methane rich vapor product of the processing column by re-converting/condensing it fully back to liquid. The processing column may be employed in various modes of operation. The processing column can function as any type of column, such as for example and not by way of limitation, as a distillation column, extractive distillation column, reboiled absorption column, absorption column, lean oil absorber column, fractionation column, stripping column, refluxed stripping column, reboiled stripping column, and the like. These columns can be emulated in various proportions effectively making this a hybrid column with various degrees of functions of these column configurations with additional pressure swing functionality or pressure variability functionality to match most effective use of this process of disclosure.

-11-

[0052] Highlights of the present invention include: unique arrangement of the pre-column, at/within-column, and post-column major parts of the process; flexible variability of the RLNG source pressure (via pump 22 or other motive power system) – this pressure control is important for the rest of the process; the heat exchanger 30 operation; regulating internal pressure of the RLNG-cold side of exchanger via a back-pressure controller (33); the feed composition/enthalpy variability options with item degasser 40; the column operability functions -- variability with multiple optional combinations of column source streams and their locations on the column, the variability of column pressure and its effect on its operations (via back pressure controller (16)); the overall combination of items downstream of stream 8 and 9 to provide variability of the pressure on the exchanger condensing hot side; and optional process embodiment to re-produce LLNG as vaporized LLNG without compression equipment with economy of equipment/complexity/energy.

[0053] A flexibility integrated methodology and system for extracting less volatile components from a fluid stream working in high percentage or low percentage extracting modes is disclosed, such as, showing the extracting/rejecting of ethane in this inventive process (of extracting NGL from LNG) while not requiring vaporization of Rich LNG (RLNG) prior to separation in to LNG/NGL in a column(s) and essentially on a large part eliminating the requirements of compressing (optional) any gas to produce Lean LNG (LLNG) while taking the feed RLNG extracting its NGLs in the processing system and reproducing Lean LNG of spec from Rich LNG.

[0054] Referring to FIGS. 1A, 1B, 2A and 2B there is shown a flow diagram of a LNG processing plant used in a HYSYS® software simulation in accordance with the present invention. The Tables set forth below contain the operational parameters used in the HYSYS® simulations of the methods of the present invention. The process flow diagram shown in the Figures is further described as follows:

[0055] It is contemplated that the major equipment involved in the process will be:

[0056] A. Feed pump 22 for Rich LNG from storage/transport location 20,

[0057] B. (Optional Rich LNG sub-cooler utilizing LNG “cold”) 13,

[0058] C. LNG Exchanger 30,

[0059] D. Degasser vessel 40 (optionally can be an inline degasser).

-12-

[0060] E. Processing Column (or fractionation column) 50, its internals (Trays or Packing) and Reboiler (collectively called “Column”),

[0061] F. Column Bottom Discharge pump,

[0062] G. Receiving Vessel 70 for the condensed Column Overhead stream 7, 9,

[0063] H. (Optionally, in any other contemplated modes of operation, compressor 80 for tail gas stream 11 from Receiving Vessel 70),

[0064] I. Discharge Pump 74 for the Lean LNG condensed stream 10, 10A,

[0065] J. Flash, Pump and distribution equipment (not shown) for product Lean LNG (stream 10A or its partial product as an option equipment to further recycle part of the Lean LNG (stream 10A) back to storage 20 as “Roll Over” control method wherever it is implemented.

[0066] More particularly, referring to the figures, an LNG (or RICH LNG) stream 1 is pumped via pump 22 from a storage location 20 (e.g., tank 20 having LNG level 20A) through suitable conduit through a valve 24, where it becomes labeled stream 1A. It will be understood by those of ordinary skill in the art that suitable conduit is used throughout the flow diagram to connect together the various components as shown to permit (as described) fluid communication between components for transporting the various product streams therein. Stream 1A is directed into the cool side of a heat exchanger 30 via heat exchanger inlet 31, and is discharged out of the heat exchanger 30 as stream 2 via heat exchanger outlet 32. As will be further described below, the heat exchanger 30 has a hot side that is in turn receiving one or more warm process stream(s), particularly the overhead stream 7 from a processing column 50. The warmed LNG stream 2, in its liquid state, is passed through a valve 33 where it becomes stream 3 or 3A. The valve 33 can be used to, e.g., regulate the pressure of stream 3 and 3A, for example, to lower the pressure of stream 3 after stream 3 exits the heat exchanger 30 prior to entering the processing column 50. In one embodiment, the warmed LNG stream 3, still in liquid state, is then passed into a degassing vessel 40 via degasser inlet 43, though no vaporization is anticipated under the normal modes of operation of the invention. The degassed liquid stream 4 is discharged, via degasser lower outlet 44 for discharging stream 4, from the vessel 40 and is fed to the processing column 50 at desired location(s) (using an optional pump 48, as may be desired). The degassed gas stream 5 is discharged, via degasser upper outlet 45, from the vessel 40 and is fed to the processing

-13-

column 50 at desired location(s) via inlet(s) 55 (and in connection with suitable valve(s) 17D). In another embodiment, the warmed LNG stream 3A, still in liquid state, is then passed directly to the processing column 50 and introduced into the column at any desired location via one or more suitable inlet ports (not shown).

[0067] The processing column 50 is outfitted with one or more processing column inlet(s) 54 for receiving stream 4 at various locations along the height of the column 50 (in connection with suitable valve(s) 17E). The column 50 is also outfitted with one or more processing column inlet(s) 55 for receiving stream 5 at various locations along the height of the column 50. In the processing column 50, the more and less volatile components are separated and the lighter components (“overhead”, stream 7) predominantly are discharged out of the column 50 from the upper section (via processing column outlet 57) and the less volatile components (“bottoms”, stream 6) are discharged from a lower section (via processing column outlet 56).

[0068] The column bottoms may be directed to storage or end use locations or be circulated to be mixed with the warmed feed stream or alternatively connected to another portion of the column (which can be piped to be able to provide inlet to the column at various stages or locations in the column for flexibility) that would optimize the NGL extraction with C2 extraction or Rejection mode operations or to obtain specified NGL requirements.

[0069] The bottoms (stream 6) from processing column 50 are discharged (via discharge port 56) to a pump (not shown) which may circulate some of the bottoms back to the column 50 or the degassing vessel 40. For example, the bottoms (stream 6) may be directed into any type of column bottoms reboiler arrangement 60 via reboiler inlet 66. The reboiler has an energy stream 60A for heat in the reboiler 60. The NGLs from stream 6 may be discharged from the reboiler 60 as NGL stream 6A via reboiler outlet 66A. NGL stream 6A may be recycled to degasser 40 (through suitable valve 17C) and introduced therein via degasser inlet 46. NGL stream 6A may also be recycled (as stream 6A-1) to the processing column 50 and introduced therein via various processing column inlet(s) 56A for receiving stream 6A-1 at various locations along the height of the column 50 in coordination with suitable valve(s) 17F. The end product NGLs from stream 6 may also be discharged from the reboiler 60 as NGL stream 6 via reboiler outlet 66B where they can be directed to a desired end use/storage location (not shown), or where they could be directed to a splitting junction / valve 85 via inlet 86. The splitter/valve 85 could direct stream 6 out the outlet 87 (as stream

-14-

6A) or out the outlet 88 (as stream 6) to a desired location (not shown). The liquid NGLs could also be boiled within the reboiler and directed, via reboiler outlet 66C as stream 6C, to the processing column 50 and introduced therein via various processing column inlet(s) 56C for receiving stream 6C at various locations along the height of the column 50 in coordination with suitable valve(s) 17H.

[0070] In the processing column 50, the lighter components (“overhead”, stream 7) predominantly leave the column 50 from the upper section (via processing column outlet 57 for discharging stream 7) and are then directed through valve 16 (used to regulate pressure) where stream 7 becomes labeled as stream 8. Alternatively, stream 7 could be directed through a compressor (not shown) whereafter stream 7 would be labeled stream 8.

[0071] The overhead stream 7, 8 is directed into the hot side of the heat exchanger 30 via heat exchanger inlet 38 where the stream 8 is cooled and condensed against the Rich LNG stream in the LNG exchanger 30. In this manner, the overhead methane rich vapor stream (7, 8) from the column 50 is diverted to the LNG exchanger 30 where it is condensed in cross exchange of heat with the cold Rich LNG feed (1A) and is anticipated to condense up to 100% into Lean LNG liquid (stream 9) which is then directed out of the exchanger 30 via exchanger outlet 39. The Lean LNG (stream 9) (which is a C1-rich mixture of liquid and gas)) is directed into and stored in a surge drum or receiving vessel 70 (via vessel inlet 71). The Lean LNG stream may be moved from the vessel 70 (as liquid stream 10) via vessel outlet 72 by way of a pump 74 at a required/desired pressure (stream 10A).

[0072] Stream 10A (Lean LNG) can then be further directed to storage or other desired location (e.g., storage tank, pressurized pipeline, not shown).

[0073] Stream 10A (Lean LNG) may also be routed through a splitter or junction / valve 75 (via inlet 76). Stream 10A may be directed through valve 75 (via outlet 77) as stream 10E to a desired location or storage facility for the Lean LNG product. For example, and referring to FIG. 1B (which illustrates detail area 1B from FIG. 1A), stream 10E (Lean LNG) may be directed to one or more further heat exchanger processing unit(s) 1C. The stream 10E would enter heat exchanger 100 and depart as stream 10F through a valve 200 or other pressure maintenance system (for maintaining the desired pressure in the heat exchanger 100 to maintain the Lean LNG in liquid form as it is heated in the heat exchanger much like with the operation of exchanger 30 described herein) where stream 10F becomes stream 10G. The heat exchanger 100 also receives desired heat transfer stream 99A (which

-15-

can be an independent hot stream or a hot side stream from another part of the process) which then transfers thermal energy and departs exchanger 100 as stream 99B (which can be directed to other locations as desired). Much like with stream 3, stream 10G is fed into a degasser or other receiving vessel 300 where the liquid phase can then be directed out as Lean LNG stream 10H and any gas can be directed out as gas stream 25A. Side stream 99C can pass through the vessel 300 (to, e.g., transfer thermal energy) and exit the vessel as side stream 99D. Gas stream 25A can then pass through a valve 700 where it becomes gas stream 25B and can then be directed to a desired location, such as, by joining it with gas product stream 12A to create gas product stream 12B. Liquid stream 10H can be then directed to a desired location, such as, an inlet stream 10H for another heat exchanger processing unit.

[0074] For example, still referring to FIG. 1B, a second heat exchanger processing unit could be tied into the first heat exchanger processing unit 1C in series fashion. For example, the liquid product stream 10H (from unit 1C) would enter another heat exchanger 400 and depart as stream 10J through a backpressure/level control valve 500 or other pressure maintenance system (for maintaining the desired pressure in the heat exchanger 400 to maintain the Lean LNG in liquid form as it is heated in the heat exchanger much like with the operation of exchanger 30 described herein) where stream 10J becomes stream 10K. The heat exchanger 400 also receives desired heat transfer stream 99E (which can be an independent stream or a side stream from another part of the process) which then transfers thermal energy and departs exchanger 400 as stream 99F (which can be directed to other locations as desired). Much like with stream 3, stream 10K is fed into another degasser or other receiving vessel 600 where the liquid Lean LNG phase can then be directed out as Lean LNG stream 10L and any gas can be directed out as stream 25C. Side stream 99G can pass through the vessel 600 (to, e.g., transfer thermal energy) and exit the vessel as side stream 99H. Gas stream 25C can then pass through a valve 800 where it becomes gas stream 25D and can then be directed to a desired location, such as, by joining it with gas product stream 12B to create gas product stream 12C. Liquid stream 10L can be then directed to a desired location, such as, an inlet stream 10L for yet another heat exchanger processing unit (not shown). As will be understood, any desired number of heat exchanger processing units may be arranged in series relationship. Further, it will be understood that Lean LNG stream 10E may be directed to one or more heat exchanger processing units that are themselves arranged in parallel fashion. It will be further understood that the various streams exiting the heat exchangers and degassers could also be directed to other equipment

-16-

combining/separating/collecting liquid LLNG and gas after the degassers, etc. As such, there is great versatility in the arrangement, configuration and number of heat exchanger processing units that may be employed. The LLNG is maintained in liquid form as it traverses through the exchanger(s) (where it is heated), and then after passing through the backpressure/level control valve, it may then vaporize where the liquid/vapour mixture is directed into the next degasser or a vessel or a collection of equipment, such as a pipe header, etc. When the liquid from the degasser is directed to the next heat exchanger, this provides a liquid stream feed to the heat exchanger which permits a more compact exchanger resulting in greater economy. The heat source for the heat exchangers can be air, as in air exchangers, or sea water, as in sea water exchangers, or other heat sources known in the art.

[0075] Referring back to FIG. 1A, stream 10A may be directed out of valve 75 (via outlet 78) as stream 10B to be recycled and introduced into the processing column 50 via processing column inlet(s) 58 for receiving stream 10B at various locations along the height of the column 50 in connection with valve(s) 17G. The processing column 50 receives, when required in its operating mode, a cold Lean LNG (stream 10B) at a point in the column calculated for a particular combination of pressures and fluid compositions to enhance its C2 Recovery or its C2 Rejection mode of operation.

[0076] Additionally, Lean LNG stream 10A could be diverted out of valve 75 via outlet 79 (as stream 10C, 10D, with assistance of pump 74A as may be necessary) back to storage tank 20 to permit the Lean LNG product of this invention to be recycled to storage as part of a “roll over” control method. For example, it is contemplated spreading a part of the Lean LNG product 10A (as streams 10C, 10D) as a further and part product of the Lean LNG flash of product of the Lean LNG product of this invention which can be recycled to storage as part of a “roll over” control method. In another embodiment, it is contemplated spreading a processed or cooled LNG product 10A, 10C within or above a stored quantity of LNG (such as is stored in tank 20) as part of a “roll over” control method via one or more jets or spargers (20B, 20C) within the tank 20. The sparged LNG product 10A, 10C can be introduced within (20B) or above (20C) a stored quantity of LNG (e.g., in tank 20 having LNG level 20A and one or more jets/spargers 20B, 20C). Also, it is contemplated sparging a processed LNG product 10A, 10C, 10D comprising of vapor and/or liquid flash of this disclosed process, within or above a stored quantity of LNG in tank 20.

-17-

[0077] Additionally, referring again to receiving vessel 70, tail gas stream 11 may be directed from receiving vessel 70, via outlet 73 into a compressor 80 (via compressor inlet 81) whereafter the compressed gas stream 12 emerges from compressor via outlet 82 and may be directed to a desired location. The compressor has an energy stream 80A for driving the compressor 80.

[0078] In one embodiment (referring to FIG. 2A, FIG. 2B), stream 12 may be directed into a heat exchanger CDX 90 via inlet 91 to cool stream 12. The Lean LNG cooler gas stream 12B (emerging from exchanger outlet 92) may be directed (as stream 12C) to be merged along with stream 10A and delivered to a desired location, or may be directed (as stream 12D) to be merged with stream 10B for recycling to the processing column 50. Optionally, the heat exchanger 90 may utilize an external refrigeration option 14, such as via coolant lines 14A, 14B to provide coolant.

[0079] In another embodiment, for example, and referring to FIG. 1B (which illustrates detail area 1B from FIG. 1A), stream 12 may be directed through valve 900 where it becomes gas product stream 12A where it is directed to a desired location, along with potential other gas product streams 12B, 12C as described above.

[0080] Additionally, a cold LNG stream 13 (or other desired cold stream, such as, a lean oil extraction / absorption stream) may be introduced into the heat exchanger 90 (via inlet 94) and directed out of the heat exchanger (via outlet 93) as stream 15. Stream 15 may be diverted (as stream 15A) into the degasser 40 (via inlet 47) in connection with suitable valve 17B. Stream 15 may also be directed (as stream 15B) (in connection with suitable valve 17A) into processing column 50 via processing column inlet(s) 59 for receiving stream 15B at various locations along the height of the column 50 (in association with valve(s) not shown). Although only one inlet 59 is shown, multiple inlets, at various locations along the length of the column could be employed to introduce stream 15B into the column. Stream 13 can be a stream, e.g. but not limited to, C1-rich or a C2-rich or a C3-rich or a C4-rich or a rich LNG or a Lean LNG which can act as the cooling colder stream used to condense any vapors in stream 12 in the heat exchanger 90 to give a partially or fully condensed stream 12B.

[0081] Stream 13 in another instances or embodiments, but not limited to, can be a what is termed a "Lean Oil" absorber stream that can be used to cool stream 12 in the Heat exchanger 90 or in another instance, bypass the exchanger 90 as stream 15, which in turn can

-18-

be another feed stream to the processing column as stream 15A or stream 15B to affect extraction of less volatile components from the VLNG/RLNG or be used to control separation behavior of products and operation of the processing column 50. Effectively it is an optional part of the inventive disclosure and embodiment, that is used to additionally or further control hydrocarbon mixture separation behaviors in the column similar to where the column bottoms (stream 6A or 6A-1) are used to alter feed composition to the column or separately directly fed to the column via various location connections in the column one instance of result as shown and demonstrated in Table 6.

[0082] An external refrigeration option 14 is provided in which may comprise any other material stream of choice which is a refrigerating/cooling stream (14A, 14B) that cools the stream 12 in the exchanger 90 producing a condensed fully liquid or partially liquid stream 12B.

[0083] For a better understanding of the operation of the present invention, reference is made to the following Tables in connection with process flow diagrams illustrated in the drawings.

[0084] Summary Tables

TABLE 1 - FIG 1							
Name	0-RICH LNG	0-LEAN LNG	0-NGL	Name	0-RICH LNG	0-LEAN LNG	0-NGL
Vapour Fraction	0.00	0.00	0.00	Vapour Fraction	0.00	0.00	0.00
Temperature [F]	-260.00	-126.79	119.04	Temperature [F]	-260.00	-124.07	146.76
Pressure [psig]	10.00	550.00	555.00	Pressure [psig]	10.00	550.00	555.00
Molar Flow [lbmole/hr]	109804.69	108087.98	1784.40	Molar Flow [lbmole/hr]	109804.69	59223.99	50642.17
C1	0.98	1.00	0.01	C1	0.53	0.97	0.01
C2	0.01	0.00	0.50	C2	0.22	0.02	0.46
C3	0.01	0.00	0.29	C3	0.14	0.00	0.29
iC4	0.00	0.00	0.05	iC4	0.03	0.00	0.06
nC4	0.00	0.00	0.05	nC4	0.05	0.00	0.10
C5's	0.00	0.00	0.02	C5's	0.01	0.00	0.03
C6+	0.00	0.00	0.03	C6+	0.02	0.00	0.04
VOLFlow (Ethane) [barrel/day]				VOLFlow (Ethane) [barrel/day]			
VOLFlow (Propane) [barrel/day]	6356.35	1197.50	5164.17	VOLFlow (Propane) [barrel/day]	142556.61	8138.67	134475.94
	3337.23	213.83	3124.46		89321.06	1059.51	88268.67
TABLE 2 - FIG 2							
Name	0-RICH LNG	0-LEAN LNG	0-NGL	Name	0-RICH LNG	0-LEAN LNG	0-NGL
Vapour Fraction	0.00	0.00	0.00	Vapour Fraction	0.00	0.00	0.00
Temperature [F]	-260.00	-124.07	146.76	Temperature [F]	-260.00	-118.37	121.74
Pressure [psig]	10.00	550.00	555.00	Pressure [psig]	10.00	600.00	605.00
Molar Flow [lbmole/hr]	109804.69	59223.99	50642.17	Molar Flow [lbmole/hr]	109804.69	98516.48	11333.33
C1	0.53	0.97	0.01	C1	0.87	0.97	0.01
C2	0.22	0.02	0.46	C2	0.09	0.03	0.64
C3	0.14	0.00	0.29	C3	0.03	0.00	0.26
iC4	0.03	0.00	0.06	iC4	0.01	0.00	0.09
nC4	0.05	0.00	0.10	nC4	0.00	0.00	0.00
C5's	0.01	0.00	0.03	C5's	0.00	0.00	0.00
C6+	0.02	0.00	0.04	C6+	0.00	0.00	0.00
VOLFlow (Ethane) [barrel/day]				VOLFlow (Ethane) [barrel/day]			
VOLFlow (Propane) [barrel/day]	142556.61	8138.67	134475.94	VOLFlow (Propane) [barrel/day]	56796.57	15135.38	41754.11
	89321.06	1059.51	88268.67		19270.43	1585.35	17709.47
TABLE 3 - FIG 3							
Name	0-RICH LNG	0-LEAN LNG	0-NGL	Name	0-RICH LNG	0-LEAN LNG	0-NGL
Vapour Fraction	0.00	0.00	0.00	Vapour Fraction	0.00	0.00	0.00
Temperature [F]	-260.00	-123.37	143.45	Temperature [F]	-260.00	-118.37	121.74
Pressure [psig]	10.00	550.00	555.00	Pressure [psig]	10.00	600.00	605.00
Molar Flow [lbmole/hr]	109804.69	43372.64	66450.93	Molar Flow [lbmole/hr]	109804.69	98516.48	11333.33
C1	0.39	0.96	0.01	C1	0.87	0.97	0.01
C2	0.32	0.03	0.51	C2	0.09	0.03	0.64
C3	0.12	0.00	0.19	C3	0.03	0.00	0.26
iC4	0.04	0.00	0.07	iC4	0.01	0.00	0.09
nC4	0.08	0.00	0.13	nC4	0.00	0.00	0.00
C5's	0.02	0.00	0.03	C5's	0.00	0.00	0.00
C6+	0.03	0.00	0.04	C6+	0.00	0.00	0.00
VOLFlow (Ethane) [barrel/day]				VOLFlow (Ethane) [barrel/day]			
VOLFlow (Propane) [barrel/day]	202278.20	7859.77	194437.13	VOLFlow (Propane) [barrel/day]	56796.57	15135.38	41754.11
	75459.14	589.61	74870.88		19270.43	1585.35	17709.47
TABLE 4 - FIG 4							
Name	0-RICH LNG	0-LEAN LNG	0-NGL	Name	0-RICH LNG	0-LEAN LNG	0-NGL
Vapour Fraction	0.00	0.00	0.00	Vapour Fraction	0.00	0.00	0.00
Temperature [F]	-260.00	-97.33	162.21	Temperature [F]	-260.00	-118.37	121.74
Pressure [psig]	10.00	550.00	555.00	Pressure [psig]	10.00	600.00	605.00
Molar Flow [lbmole/hr]	109804.69	60407.94	53482.33	Molar Flow [lbmole/hr]	109804.69	98516.48	11333.33
C1	0.39	0.70	0.01	C1	0.87	0.97	0.01
C2	0.32	0.26	0.42	C2	0.09	0.03	0.64
C3	0.12	0.03	0.21	C3	0.03	0.00	0.26
iC4	0.04	0.01	0.08	iC4	0.01	0.00	0.09
nC4	0.08	0.01	0.15	nC4	0.00	0.00	0.00
C5's	0.02	0.00	0.04	C5's	0.00	0.00	0.00
C6+	0.03	0.00	0.05	C6+	0.00	0.00	0.00
VOLFlow (Ethane) [barrel/day]				VOLFlow (Ethane) [barrel/day]			
VOLFlow (Propane) [barrel/day]	202278.20	89593.20	131357.21	VOLFlow (Propane) [barrel/day]	56796.57	15135.38	41754.11
	75459.14	11210.88	67118.70		19270.43	1585.35	17709.47
TABLE 5 - FIG 5 - FAILED CASE							
Name	0-RICH LNG	0-LEAN LNG	0-NGL	Name	0-RICH LNG	0-LEAN LNG	0-NGL
Vapour Fraction	0.00	0.00	0.00	Vapour Fraction	0.00	0.00	0.00
Temperature [F]	-260.00			Temperature [F]	-260.00		
Pressure [psig]	10.00	550.00	555.00	Pressure [psig]	10.00		
Molar Flow [lbmole/hr]	109804.69			Molar Flow [lbmole/hr]	109804.69		
C1	0.87			C1	0.87		
C2	0.09			C2	0.09		
C3	0.03			C3	0.03		
iC4	0.01			iC4	0.01		
nC4	0.00			nC4	0.00		
C5's	0.00			C5's	0.00		
C6+	0.00			C6+	0.00		
VOLFlow (Ethane) [barrel/day]				VOLFlow (Ethane) [barrel/day]			
VOLFlow (Propane) [barrel/day]	56796.57			VOLFlow (Propane) [barrel/day]	56796.57		
	19270.43				19270.43		
TABLE 6 - FIG 6 - READJUSTED CASE							
Name	0-RICH LNG	0-LEAN LNG	0-NGL	Name	0-RICH LNG	0-LEAN LNG	0-NGL
Vapour Fraction	0.00	0.00	0.00	Vapour Fraction	0.00	0.00	0.00
Temperature [F]	-260.00	-118.37	121.74	Temperature [F]	-260.00	-118.37	121.74
Pressure [psig]	10.00	600.00	605.00	Pressure [psig]	10.00	600.00	605.00
Molar Flow [lbmole/hr]	109804.69	98516.48	11333.33	Molar Flow [lbmole/hr]	109804.69	98516.48	11333.33
C1	0.87	0.97	0.01	C1	0.87	0.97	0.01
C2	0.09	0.03	0.64	C2	0.09	0.03	0.64
C3	0.03	0.00	0.26	C3	0.03	0.00	0.26
iC4	0.01	0.00	0.09	iC4	0.01	0.00	0.09
nC4	0.00	0.00	0.00	nC4	0.00	0.00	0.00
C5's	0.00	0.00	0.00	C5's	0.00	0.00	0.00
C6+	0.00	0.00	0.00	C6+	0.00	0.00	0.00
VOLFlow (Ethane) [barrel/day]				VOLFlow (Ethane) [barrel/day]			
VOLFlow (Propane) [barrel/day]	56796.57	15135.38	41754.11	VOLFlow (Propane) [barrel/day]	56796.57	15135.38	41754.11
	19270.43	1585.35	17709.47		19270.43	1585.35	17709.47

-20-

[0085] Referring to the top row of each Table, the Table number and Figure number are referenced, and the Stream labels on the second row are shown. The 3 streams shown in this Summary of Tables above are the Feed Stream (“0-RICH LNG” also referred to herein as “RICH LNG”) and Lean LNG Product Stream (“0-LEAN LNG” or “LEAN LNG”) and the NGL Product stream (“0-NGL” or “NGL”) results.

[0086] Key: (Description = Stream Name) ::

[0087] C2+ rich feed Rich LNG = 0-RICH-LNG (Liquid)

[0088] C1 rich lean LNG product = 0-LEAN LNG

[0089] C2+ rich NGL product = 0-NGL

[0090] As seen from the analysis of stream “0-RICH LNG” of Table 1 through to Table 6, the invention has been designed to handle C2 compositions of LNG of from about +/- 1% stored at about atmospheric and more particularly approximately 10 PSIG and -260°F to a C2+ content range that extends even beyond 32+% (shown) and (not shown) beyond even 52% Mol of C2 stored at 10 PSIG and -232°F.

[0091] DETAILED TABLES for all relevant streams:

[0092] As a means of the explanation of the Figures, Tables 1A through 6A are provided giving more detailed data description of the parameters for the design and operation of the process plant. It will be apparent to one skilled in the art having the benefit of the present disclosure, that the present invention could be practiced by following the present disclosure of the diagrams/figures and the accompanying data Tables. The current disclosure is indicative of reasonable assumptions typically made by those skilled in the art, including rounding of the data, ambient conditions and heat losses not accounted and not shown but contemplated where required.

[0093] Key:

[0094] The rows showing label “Name” is for Stream Labels in that row and which are directly referenced to stream data from the FIG. 1A process flow diagram.

[0095] First row indicates the Table Number.

TABLE 1A

TABLE 1A - FIG 1											
Name	0-RICH LNG	0-LEAN LNG	0-NGL	1	1A	2	3	4	5		
Vapour Fraction	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	1.0000		
Temperature [F]	-260.0	-126.8	119.0	-260.0	-260.0	-162.4	-162.8	-162.8	-162.8		
Pressure [psig]	10.0	550.0	555.0	700.0	700.0	693.0	560.0	555.0	555.0		
Molar Flow [lbmole/hr]	109,804.7	108,088.0	1,784.4	109,804.7	109,804.7	109,804.7	109,804.7	109,804.7	-		
C1	0.9800	0.9960	0.0081	0.9800	0.9800	0.9800	0.9800	0.9800	0.9998		
C2	0.0100	0.0019	0.4999	0.0100	0.0100	0.0100	0.0100	0.0100	0.0002		
C3	0.0051	0.0003	0.2938	0.0051	0.0051	0.0051	0.0051	0.0051	0.0000		
iC4	0.0008	0.0000	0.0477	0.0008	0.0008	0.0008	0.0008	0.0008	0.0000		
nC4	0.0009	0.0000	0.0541	0.0009	0.0009	0.0009	0.0009	0.0009	0.0000		
C5s	0.0003	0.0000	0.0182	0.0003	0.0003	0.0003	0.0003	0.0003	0.0000		
C6+	0.0005	0.0000	0.0305	0.0005	0.0005	0.0005	0.0005	0.0005	0.0000		
VOLFlow (Ethane) [barrel/day]	6,356.35	1,197.50	5,164.17	6,356.35	6,356.35	6,356.35	6,356.35	6,356.35	-		
VOLFlow (Propane) [barrel/day]	3,337.23	213.83	3,124.46	3,337.23	3,337.23	3,337.23	3,337.23	3,337.23	-		
Name	6	6A	7	8	9	10	10B	11	12		
Vapour Fraction	0.0000	0.0000	1.0000	1.0000	0.0000	0.0000	0.0000	1.0000	<empty>		
Temperature [F]	119.0	119.0	-125.0	-125.0	-127.1	-127.1	-126.8	-127.1	<empty>		
Pressure [psig]	555.0	555.0	550.0	550.0	543.0	543.0	550.0	543.0	555.0		
Molar Flow [lbmole/hr]	1,784.4	-	120,030.1	120,097.8	120,097.8	120,097.8	12,009.8	-	-		
C1	0.0081	0.0081	0.9961	0.9960	0.9960	0.9960	0.9960	0.9986	0.9986		
C2	0.4999	0.5002	0.0019	0.0019	0.0019	0.0019	0.0019	0.0006	0.0006		
C3	0.2938	0.2937	0.0003	0.0003	0.0003	0.0003	0.0003	0.0000	0.0000		
iC4	0.0477	0.0477	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000		
nC4	0.0541	0.0540	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000		
C5s	0.0182	0.0182	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000		
C6+	0.0305	0.0305	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000		
VOLFlow (Ethane) [barrel/day]	5,164.17	-	1,325.24	1,330.55	1,330.55	1,330.55	133.06	-	-		
VOLFlow (Propane) [barrel/day]	3,124.46	-	236.53	237.59	237.59	237.59	23.76	-	-		

-22-

[0097] TABLE 1A (in conjunction with the process flow diagram of FIG. 1A) shows the processing of LNG with 1% C2 in an Ethane Recovery mode, resulting in a recovery of 81% of C2 and 94% of C3, with the rest of the component recoveries reflected in Tables 1 and 1A. Regarding the data in TABLE 1A, (and in reference to FIG. 1A) it is noted that stream 2 is maintained as a liquid with minimal reflux (stream 10B, with < than 1% C2 - as same as C2 in Lean LNG) and 0 bottoms recycle (stream 6A). No compression/recompression is required for tail stream 11 which can be tied into a gasification system and pipeline much more economically at a higher pressure with just some addition of heat and compression to even higher pressure if desired. For example, referring to streams 1A, 2 and 3, the vapour fraction is zero thereby indicating that there is no vaporization of the LNG stream. Stream 5 reflects a “default” stream/pipe for vapor to leave if any gas degasses – so always a Vapour Fraction of “1” - a simulation stream vapor and liquid from any vessel. However, there is no “flow quantity” in stream 5 – Molar Flow – the dash (–) means zero.

[0098] TABLE 2A

TABLE 2A - FIG 2											
Name	0-RICH LING	0-LEAN LING	0-NGL	1	1A	2	3	4	5		
Vapour Fraction	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	1.0000		
Temperature [F]	-260.0	-124.1	146.8	-260.0	-260.0	-187.2	-186.4	-186.4	-186.4		
Pressure [psig]	10.0	550.0	555.0	700.0	700.0	693.0	560.0	555.0	555.0		
Molar Flow [lbmole/hr]	109,804.7	59,224.0	50,642.2	109,804.7	109,804.7	109,804.7	109,804.7	109,804.7	-		
C1	0.5279	0.9723	0.0084	0.5279	0.5279	0.5279	0.5279	0.5279	0.5278		
C2	0.2243	0.0237	0.4587	0.2243	0.2243	0.2243	0.2243	0.2243	0.2243		
C3	0.1365	0.0030	0.2925	0.1365	0.1365	0.1365	0.1365	0.1365	0.1365		
iC4	0.0265	0.0002	0.0573	0.0265	0.0265	0.0265	0.0265	0.0265	0.0265		
nC4	0.0475	0.0002	0.1028	0.0475	0.0475	0.0475	0.0475	0.0475	0.0475		
C5's	0.0122	0.0000	0.0265	0.0122	0.0122	0.0122	0.0122	0.0122	0.0122		
C6+	0.0166	0.0000	0.0359	0.0166	0.0166	0.0166	0.0166	0.0166	0.0166		
VOLFlow (Ethane) [barrel/day]	142,556.61	8,138.67	134,475.94	142,556.61	142,556.61	142,556.61	142,556.61	142,556.61	-		
VOLFlow (Propane) [barrel/day]	89,321.06	1,059.51	88,268.67	89,321.06	89,321.06	89,321.06	89,321.06	89,321.06	-		
Name	6	6A	7	8	9	10	10B	11	12		
Vapour Fraction	0.0000	0.0000	1.0000	1.0000	0.0000	0.0000	0.0000	1.0000	<empty>		
Temperature [F]	146.8	146.8	-102.9	-102.8	-124.3	-124.3	-124.1	-124.3	<empty>		
Pressure [psig]	555.0	555.0	550.0	550.0	543.0	543.0	550.0	543.0	555.0		
Molar Flow [lbmole/hr]	50,642.2	-	65,743.0	65,804.4	65,804.4	65,804.4	6,580.4	-	-		
C1	0.0084	0.0084	0.9725	0.9723	0.9723	0.9723	0.9723	0.9930	0.9930		
C2	0.4587	0.4583	0.0236	0.0237	0.0237	0.0237	0.0237	0.0062	0.0062		
C3	0.2925	0.2927	0.0030	0.0030	0.0030	0.0030	0.0030	0.0003	0.0003		
iC4	0.0573	0.0573	0.0002	0.0002	0.0002	0.0002	0.0002	0.0000	0.0000		
nC4	0.1028	0.1029	0.0002	0.0002	0.0002	0.0002	0.0002	0.0000	0.0000		
C5's	0.0265	0.0265	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000		
C6+	0.0359	0.0359	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000		
VOLFlow (Ethane) [barrel/day]	134,475.94	-	8,984.97	9,042.97	9,042.97	9,042.97	904.30	-	-		
VOLFlow (Propane) [barrel/day]	88,268.67	-	1,170.11	1,177.23	1,177.23	1,177.23	117.72	-	-		

-24-

[0099] TABLE 2A (in conjunction with the process flow diagram of FIG. 1A) shows the processing of LNG with 22% C2 in an Ethane Recovery mode, resulting in a recovery of 94% of C2 and 99% of C3, with the rest of the component recoveries reflected in Tables 2 and 2A. Regarding the data in TABLE 2A (and in reference to FIG. 1A), it is noted that steam 2 is maintained as liquid with minimal reflux (stream 10B, with > than 2% C2 - as same as C2 in Lean LNG) and 0 bottoms recycle (stream 6A). No compression/recompression is required for stream 11 – which can be tied into a gasification system and pipeline much more economically at a higher pressure with just some addition of heat and compression to even higher pressure if desired.

-25-

TABLE 3A

TABLE 3A - FIG 3											
Name	0-RICH LING	0-LEAN LING	0-NGL	1	1A	2	3	4	5		
Vapour Fraction	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	1.0000		
Temperature [F]	-260.0	-123.4	143.4	-260.0	-260.0	-206.0	-205.1	-205.0	-205.0		
Pressure [psig]	10.0	550.0	555.0	700.0	700.0	693.0	560.0	555.0	555.0		
Molar Flow [lbmole/hr]	109,804.7	43,372.6	66,450.9	109,804.7	109,804.7	109,804.7	109,804.7	109,804.7	-		
C1	0.3861	0.9649	0.0085	0.3861	0.3861	0.3861	0.3861	0.3861	0.3861		
C2	0.3182	0.0313	0.5055	0.3182	0.3182	0.3182	0.3182	0.3182	0.3182		
C3	0.1153	0.0023	0.1891	0.1153	0.1153	0.1153	0.1153	0.1153	0.1153		
iC4	0.0430	0.0003	0.0708	0.0430	0.0430	0.0430	0.0430	0.0430	0.0430		
nC4	0.0770	0.0003	0.1270	0.0770	0.0770	0.0770	0.0770	0.0770	0.0770		
C5's	0.0198	0.0000	0.0327	0.0198	0.0198	0.0198	0.0198	0.0198	0.0198		
C6+	0.0268	0.0000	0.0443	0.0268	0.0268	0.0268	0.0268	0.0268	0.0268		
VOLFlow (Ethane) [barrel/day]	202,278.20	7,859.77	194,437.13	202,278.20	202,278.20	202,278.20	202,278.20	202,278.20	-		
VOLFlow (Propane) [barrel/day]	75,459.14	589.61	74,870.88	75,459.14	75,459.14	75,459.14	75,459.14	75,459.14	-		
Name	6	6A	7	8	9	10	10B	11	12		
Vapour Fraction	0.0000	0.0000	1.0000	1.0000	0.0000	0.0000	0.0000	1.0000	<empty>		
Temperature [F]	143.4	143.6	-98.3	-98.2	-123.6	-123.6	-123.4	-123.6	<empty>		
Pressure [psig]	555.0	555.0	550.0	550.0	543.0	543.0	550.0	543.0	555.0		
Molar Flow [lbmole/hr]	66,450.9	-	48,172.9	48,191.8	48,191.8	48,191.8	4,819.2	-	-		
C1	0.0085	0.0085	0.9649	0.9649	0.9649	0.9649	0.9649	0.9906	0.9906		
C2	0.5055	0.5049	0.0312	0.0313	0.0313	0.0313	0.0313	0.0079	0.0079		
C3	0.1891	0.1893	0.0023	0.0023	0.0023	0.0023	0.0023	0.0002	0.0002		
iC4	0.0708	0.0709	0.0003	0.0003	0.0003	0.0003	0.0003	0.0000	0.0000		
nC4	0.1270	0.1271	0.0003	0.0003	0.0003	0.0003	0.0003	0.0000	0.0000		
C5's	0.0327	0.0327	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000		
C6+	0.0443	0.0444	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000		
VOLFlow (Ethane) [barrel/day]	194,437.13	-	8,714.37	8,733.08	8,733.08	8,733.08	873.31	-	-		
VOLFlow (Propane) [barrel/day]	74,870.88	-	653.77	655.12	655.12	655.12	65.51	-	-		

-26-

[00101] TABLE 3A (in conjunction with the process flow diagram of FIG. 1A) shows the processing of LNG with 32% C2 in an Ethane Recovery mode, resulting in a recovery of 96% of C2 and 99% of C3, with the rest of the component recoveries reflected in Tables 3 and 3A. Regarding the data in TABLE 3A (and in reference to FIG. 1A), it is noted that stream 2 is maintained as a liquid with minimal reflux (stream 10B, with > than 3% C2 - as same as C2 in Lean LNG) and 0 bottoms recycle (stream 6A). No compression/recompression required for stream 11 – which can be tied into a gasification system and pipeline much more economically at a higher pressure with just some addition of heat and compression to even higher pressure if desired.

-27-

[00102] TABLE 4A

TABLE 4A - FIG 4											
Name	0-RICH LING	0-LEAN LING	0-NGL	1	1A	2	3	4	5		
Vapour Fraction	0.0000	0.0000	0.0001	0.0000	0.0000	0.0000	0.0053	0.0000	1.0000		
Temperature [F]	-260.0	-97.3	162.2	-260.0	-260.0	-45.0	-45.3	30.0	30.0		
Pressure [psig]	10.0	550.0	555.0	700.0	700.0	693.0	560.0	555.0	555.0		
Molar Flow [lbmole/hr]	109,804.7	60,407.9	53,482.3	109,804.7	109,804.7	109,804.7	109,804.7	141,879.5	21,413.8		
C1	0.3861	0.6961	0.0086	0.3861	0.3861	0.3861	0.3861	0.2011	0.6692		
C2	0.3182	0.2562	0.4243	0.3182	0.3182	0.3182	0.3182	0.3656	0.2697		
C3	0.1153	0.0311	0.2106	0.1153	0.1153	0.1153	0.1153	0.1626	0.0397		
iC4	0.0430	0.0053	0.0840	0.0430	0.0430	0.0430	0.0430	0.0638	0.0072		
nC4	0.0770	0.0069	0.1526	0.0770	0.0770	0.0770	0.0770	0.1156	0.0095		
C5's	0.0198	0.0008	0.0400	0.0198	0.0198	0.0198	0.0198	0.0302	0.0012		
C6+	0.0268	0.0008	0.0545	0.0268	0.0268	0.0268	0.0268	0.0411	0.0012		
VOLFlow [Ethane] [barrel/day]	202,278.20	89,593.20	131,357.21	202,278.20	202,278.20	202,278.20	202,278.20	300,230.85	33,437.23		
VOLFlow [Propane] [barrel/day]	75,459.14	11,210.88	67,118.70	75,459.14	75,459.14	75,459.14	75,459.14	137,505.77	5,070.95		
Name	6	6A	7	8	9	10	10B	11	12		
Vapour Fraction	0.0001	0.0000	1.0000	1.0000	0.0000	0.0000	0.0000	1.0000	<empty>		
Temperature [F]	162.2	162.2	9.1	16.7	-97.4	-97.4	-97.3	-97.4	<empty>		
Pressure [psig]	555.0	555.0	550.0	550.0	543.0	543.0	550.0	543.0	555.0		
Molar Flow [lbmole/hr]	106,964.7	53,488.6	96,600.6	100,679.9	100,679.9	100,679.9	40,272.0	-	-		
C1	0.0086	0.0086	0.7243	0.6961	0.6961	0.6961	0.6961	0.9445	0.9445		
C2	0.4243	0.4243	0.2337	0.2562	0.2562	0.2562	0.2562	0.0513	0.0513		
C3	0.2106	0.2106	0.0275	0.0311	0.0311	0.0311	0.0311	0.0015	0.0015		
iC4	0.0840	0.0840	0.0045	0.0053	0.0053	0.0053	0.0053	0.0001	0.0001		
nC4	0.1526	0.1526	0.0059	0.0069	0.0069	0.0069	0.0069	0.0001	0.0001		
C5's	0.0400	0.0400	0.0007	0.0008	0.0008	0.0008	0.0008	0.0000	0.0000		
C6+	0.0545	0.0545	0.0007	0.0008	0.0008	0.0008	0.0008	0.0000	0.0000		
VOLFlow [Ethane] [barrel/day]	262,714.43	131,389.87	130,682.44	149,321.99	149,321.99	149,321.99	59,728.80	-	-		
VOLFlow [Propane] [barrel/day]	134,237.40	67,117.57	15,813.24	18,684.80	18,684.80	18,684.80	7,473.92	-	-		

-28-

[00103] TABLE 4A (in conjunction with the process flow diagram of FIG. 1A) shows the processing of LNG with 32% C2 in an Ethane Controlled Rejection mode, resulting in a recovery of 65% of C2 and 89% of C3, with the rest of the component recoveries reflected Tables 4 and 4A. Regarding the data in TABLE 4A (and in reference to FIG. 1A), it is noted that stream 2 is maintained as a liquid with minimal reflux (stream 10B, with > than 25% C2 - as same as C2 in Lean LNG) and a particular bottoms recycle (stream 6A). No compression/recompression required for stream 11 – which can be tied into a gasification system and pipeline much more economically at a higher pressure with just some addition of heat and compression to even higher pressure if desired. TABLE 4A reflects the addition of a recycle stream 6A resulting in a changing of enthalpy/composition. This results in some vapor “quantity” that flows out as stream 5 (which is a vapor stream by default from a “separator/vessel” in simulations and is always Vapour Fraction and so a “1” and the other side Stream 4 by default is always a liquid = “0” Vapour Fraction.

[00104] TABLE 5A

TABLE 5A - FIG 5											
Name	0-RICH LNG	0-LEAN LNG	0-NGL	1	1A	2	3	4	5		
Vapour Fraction	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.1305	0.0000	1.0000		
Temperature [F]	-260.0	-124.1	111.3	-260.0	-260.0	-102.0	-111.4	-97.7	-97.7		
Pressure [psig]	10.0	550.0	555.0	700.0	700.0	693.0	560.0	555.0	555.0		
Molar Flow [lbmole/hr]	109,804.7				109,804.7						
C1	0.8710				0.8710						
C2	0.0894				0.0894						
C3	0.0294				0.0294						
iC4	0.0100				0.0100						
nC4	-				-						
C5's	-				-						
C6+	-				-						
VOLFlow (Ethane) [barrel/day]	56,796.57				56,796.57						
VOLFlow (Propane) [barrel/day]	19,270.43				19,270.43						
Name	6	6A	7	8	9	10	10B	11	12		
Vapour Fraction	0.0000	0.0000	1.0000	1.0000	0.0000	0.0000	0.0000	1.0000	<empty>		
Temperature [F]	111.3	111.2	-108.8	-108.6	-124.3	-124.3	-124.1	-124.3	<empty>		
Pressure [psig]	555.0	555.0	550.0	550.0	543.0	543.0	550.0	543.0	555.0		
Molar Flow [lbmole/hr]											
C1											
C2											
C3											
iC4											
nC4											
C5's											
C6+											
VOLFlow (Ethane) [barrel/day]											
VOLFlow (Propane) [barrel/day]											

-30-

[00105] TABLE 5A (in conjunction with the process flow diagram of FIG. 1A) shows the processing of LNG with 8.9% C2 in the same operating mode as TABLE 4/4A. It fails (as seen in TABLE 5/5A) and cannot perform unless the inventive design changes in the operating mode are made as in TABLE 6/6A. With respect to the data in TABLE 5A and 6A and in reference to the figures – it shows that for one of various pressure and temperature for stream 2, when permitted or allowed to vaporize in stream 2 TABLE 5A, the system fails and when maintained as a liquid as in stream 2 of TABLE 6A, the system along with all the other parameters performs. TABLE 5A demonstrates that control of column pressure drives the feasibility of the process as well. TABLE 5A demonstrates the effect of changing the column pressure slightly from 605 psig to 555 psig. The failure translates down to the exchanger which goes into a “temperature cross” or an impractical or uneconomic exchanger design. Stream 2 is liquid until let down in pressure at the valve down to stream 3 (exchanger pressure Stream 2 was 693 psig and valve let it down as stream 3 to 560 psig – partly vaporizing).

-31-

TABLE 6A

TABLE 6A - FIG 6											
Name	0-RICH LING	0-LEAN LING	0-NGL	1	1A	2	3	4	5		
Vapour Fraction	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	1.0000		
Temperature [F]	-260.0	-118.4	121.7	-260.0	-260.0	-114.8	-115.6	-96.3	-96.3		
Pressure [psig]	10.0	600.0	605.0	700.0	700.0	693.0	605.0	605.0	605.0		
Molar Flow [lbmole/hr]	109,804.7	98,516.5	11,332.1	109,804.7	109,804.7	109,804.7	109,804.7	100,028.8	21,137.1		
C1	0.8710	0.9702	0.0081	0.8710	0.8710	0.8710	0.8710	0.7543	0.9594		
C2	0.0894	0.0265	0.6364	0.0894	0.0894	0.0894	0.0894	0.1628	0.0359		
C3	0.0294	0.0027	0.2622	0.0294	0.0294	0.0294	0.0294	0.0613	0.0037		
IC4	0.0100	0.0004	0.0933	0.0100	0.0100	0.0100	0.0100	0.0214	0.0005		
nC4	-	0.0000	0.0000	-	-	-	-	0.0000	0.0000		
C5s	-	0.0000	0.0000	-	-	-	-	0.0000	0.0000		
C6+	-	0.0000	0.0000	-	-	-	-	0.0000	0.0000		
VOLFlow (Ethane) [barrel/day]	56,796.57	15,135.38	41,748.37	56,796.57	56,796.57	56,796.57	56,796.57	94,285.31	4,394.97		
VOLFlow (Propane) [barrel/day]	19,270.43	1,585.35	17,707.62	19,270.43	19,270.43	19,270.43	19,270.43	36,538.92	467.90		
Name	6	6A	7	8	9	10	10B	11	12		
Vapour Fraction	0.0000	0.0000	1.0000	1.0000	0.0000	0.0000	0.0000	1.0000	<empty>		
Temperature [F]	121.7	121.7	-104.0	-104.0	-118.6	-118.6	-118.4	-118.6	<empty>		
Pressure [psig]	605.0	605.0	600.0	600.0	593.0	593.0	600.0	593.0	605.0		
Molar Flow [lbmole/hr]	22,664.3	11,361.2	164,179.2	164,194.1	164,194.1	164,194.1	65,677.7	-	-		
C1	0.0081	0.0081	0.9701	0.9702	0.9702	0.9702	0.9702	0.9904	0.9904		
C2	0.6364	0.6368	0.0266	0.0265	0.0265	0.0265	0.0265	0.0088	0.0088		
C3	0.2622	0.2620	0.0027	0.0027	0.0027	0.0027	0.0027	0.0004	0.0004		
IC4	0.0933	0.0931	0.0004	0.0004	0.0004	0.0004	0.0004	0.0000	0.0000		
nC4	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000		
C5s	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000		
C6+	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000		
VOLFlow (Ethane) [barrel/day]	83,496.74	41,883.71	25,273.80	25,225.64	25,225.64	25,225.64	10,090.25	-	-		
VOLFlow (Propane) [barrel/day]	35,415.24	17,736.39	2,648.48	2,642.25	2,642.25	2,642.25	1,056.90	-	-		

[00107] TABLE 6A (in conjunction with the process flow diagram of FIG. 1A) shows It shows the processing of LNG with 8.9% C2 in an Ethane Controlled Rejection mode. With respect to the data in TABLE 6A (and in reference to FIG. 1A), it is noted that stream 2 is maintained as a liquid with some reflux (stream 10B, with > than 2% C2 - as same as C2 in Lean LNG) and 0 bottoms recycle (stream 6A). No compression/recompression is required for stream 11 – which can be tied into a gasification system and pipeline much more economically at a higher pressure with just some addition of heat and compression to even higher pressure if desired. One of the practical aspects of the invention is, and it being one subject of the invention, that the pressure has to be changed for the Column as shown (FIG. 1A, TABLES 6/6A) for the system to be able to perform as expected, with other choice of parameters essentially kept the same, resulting in a recovery of 74% of C2 and 92% of C3, with the rest of the component recoveries reflected in TABLES 6 and 6A. In TABLE 6A, although streams 4 and 5 are shown at pressure of 560 psig it is normally boosted to whatever pressure the column requires for its operation (which should have been 605psig in this case TABLE 6A – going to processing column at 600psig). This is accomplished “intrinsically” in the simulation at the column, but it could have been shown explicitly (as nominally shown for the TABLE 5A (555 psig going to column operating at 550 psig).

[00108] In the present invention, it is desired to have the pressure at stream 2 sufficiently high so there is no need for boosting of pressure mentioned, as at the valve there is stream 2 @ 693psig and the valve will control the pressure drop from stream 3 pressure to provide sufficient pressure for streams 4 & 5 – nominally at 605 psig - to supply flow to the processing column at pressure 600psig.

[00109] As noted above, FIG. 2A is another flow diagram of a HYSYS Simulation of a LNG processing plant in accordance with the present invention. It illustrates the employment of additional processing options to those outlined in connection with FIG. 1A and Tables 1-6. Specifically, FIG. 2A adds processing options related to introduction of a cold LNG or other cold desired stream into stream 12 as described above. FIG. 2B provides an enlarged view of detailed area 2B of FIG. 2A (but has relevance also to the teachings of FIG. 1A). FIG. 2B illustrates a detail of the processing column area and exemplary optional connections thereon for receiving various streams.

[00110] The current process parameters adopted for this disclosure are traditional British units. Though not shown for this disclosure the use of SI units is anticipated and contemplated where required.

[00111] For convenience, the Tables reflect all Stream Flows, Temperatures, Vapor Fractions, Compositions (shown in Mole Fractions), and further including volumetric flows of the two key components, C2 and C3, which are used to evaluate Recovery Performances. The typical parameters of measurement are indicated within the Tables.

[00112] The present invention permits flexibility in optimizing the processing column operation in its modes of operation. In Ethane (C2) Rejection mode, the processing column is optimized to produce more methane C1 as an overhead stream (and less C2+ in the bottom stream). In Ethane Extraction mode, the processing column is optimized to produce more ethane+ (C2+) fractions from the bottoms stream. The use of reflux stream 10B permits reintroduction of methane rich liquid to the processing column as a reflux to optimize the processing column, to increase or manipulate the purity of the methane overhead stream 7, or to increase or control the C2+ fraction in the bottom stream 6. The present invention permits many points of flexibility and control for operational optimization. The processing column may be optimized (via, e.g., reflux, bottoms recycle) to shift the C2 fraction in the column to exit the column as part of the overhead stream (C2 rejection mode) or as part of the bottoms stream (C2 extraction mode). A typical NGL specification seeks to remove C1 content to less than 1-10%. With the present system, the C1 content of the NGL stream can be reduced to less than 0.5%. System pressure plays a key role in this process as is outlined in the TABLES.

[00113] It is contemplated a process for separating less volatile compounds from more volatile compounds, more particularly hydrocarbons, for example the compounds less volatile than methane, in an LNG stream (herein called "Rich LNG").

[00114] It is contemplated to take an available normally low pressure cryogenic feed LNG and pump it to a higher pressure.

[00115] It is contemplated the higher pressure LNG stream 1A is pumped to a LNG Exchanger 30 to cross exchange heat with other stream(s) (e.g., stream 8).

-34-

[00116] It is contemplated the cold stream is the Rich LNG (streams 1, 1A) and the warm stream is any of the streams from further downstream process(es), more particularly in this instance the overhead vapor stream 7 from the processing column 50.

[00117] It is contemplated the Rich LNG stream 1A is heated in the LNG exchanger 30 while maintaining it in its liquid phase and state.

[00118] It is contemplated the LNG exchanger 30 to be of any particular suitable design or network of exchangers.

[00119] It is contemplated feeding the warmed Rich LNG stream 3 to a degassing vessel 40.

[00120] It is contemplated any vapor (stream 5) from the degasser vessel 40 be passed to the processing column 50.

[00121] It is contemplated the liquid (stream 4) from the degasser 40 is passed to an upper section of the processing column 50.

[00122] It is contemplated a draw is taken from the processing column bottoms or bottom stream 6 is passed to the degasser (stream 6A) or a point in the column (stream 6A-1).

[00123] It is contemplated the degasser 40 can be a liquid full vessel where no vapor or gas is evolved.

[00124] It is contemplated the degasser 40 can be of a simplest of device to whatever is dictated by the conditions.

[00125] It is completed the processing column 50 is effectively a reboiled absorber complete with the processing column 50 and a reboiler 60.

[00126] It is contemplated a stream (10B) is drawn from the Lean LNG and returned to the processing column 50.

[00127] It is contemplated the draws from the two draws, one from Lean LNG and one from the processing column bottom streams are available to work in tandem in any amounts and combinations at their inlet locations to the column.

[00128] It is contemplated the plant is operated in a way to manage the Temperature Approach within the LNG exchanger 30 so that the whole arrangement can perform practically, economically, and reasonably.

-35-

[00129] It is contemplated the plant is operated in a way to manage the LMTD (Log Mean Temperature Difference) within the LNG exchanger 30 so that the whole arrangement can perform practically, economically, and reasonably.

[00130] It is contemplated controlling the processing column 50 pressure to effectively control the LNG exchanger 30 functions.

[00131] It is contemplated controlling the LNG exchanger 30 to function practically by controlling the processing column 50 pressure.

[00132] It is contemplated controlling the LNG exchanger 30 to function practically by controlling the processing column pressure and Rich LNG feed pressure.

[00133] It is contemplated controlling the LNG exchanger to function practically by controlling the processing column 50 pressure and/or Rich LNG feed pressure and/or streams to/from the column 50.

[00134] It is contemplated controlling the LNG exchanger to function practically by controlling the Rich LNG feed pressure.

[00135] It is contemplated controlling the LNG exchanger function and/or product stream compositions and/or separations of the components received in the system, more particularly components of feed stream of Rich LNG, practically by controlling the processing column pressure and/or Rich LNG feed pressure and/or streams and their properties and their inlet/outlet locations to/from the column.

[00136] It is contemplated controlling the processing column 50 pressure to effectively control the LNG exchanger operation functions and the overall function of the whole interdependent system.

[00137] It is contemplated managing the Rich LNG feed pressure to manage the various operations and operability of the system to perform.

[00138] It is contemplated controlling temperature of stream(s) to or from the LNG exchanger either directly or indirectly.

[00139] The HYSYS run data tables included here provide example description of the system behavior and operation. The HYSYS run data tables and figures, taken together, provide substantial description of the system to enable one to design a system to operate in practice.

-36-

[00140] It is contemplated withdrawing the methane rich stream as the vapor stream from an upper section of the processing column 50.

[00141] It is contemplated sending the vapor stream 7 to the LNG exchanger 30 for condensation.

[00142] It is contemplated the less volatile than methane liquid stream mix is withdrawn from the lower section of the processing column 50.

[00143] It is contemplated a part of the Lean LNG product as a further and part product of the Lean LNG flash of product of the Lean LNG product of this invention can be recycled to storage as part of a “roll over” control method.

[00144] It is contemplated introducing a technique and method for control of storage LNG “Roll Overs”.

[00145] It is contemplated spreading a part of the Lean LNG product as a further and part product of the Lean LNG flash of product of the Lean LNG product of this invention which can be recycled to storage as part of a “roll over” control method.

[00146] It is contemplated spreading a processed or cooled LNG product within or above a stored quantity of LNG as part of a “roll over” control method via jets or spargers.

[00147] It is contemplated sparging a processed product of this disclosed process, within or above a stored quantity of LNG as part of a “roll over” control method via jets or spargers.

[00148] It is contemplated sparging a processed LNG product comprising of vapor and/or liquid flash of this disclosed process, within or above a stored quantity of LNG as part of a “roll over” control method via jets or spargers.

[00149] In a more narrative form to elucidate further the separation of Rich LNG into methane rich Lean LNG and methane depleted NGL product the present disclosure describes a process in this example more particularly described for separating and recovering ethane and heavier hydrocarbons from LNG, and could be applied to streams of other slates of hydrocarbons or non-hydrocarbons or their mixes.

[00150] The present invention includes an option to include compression.

[00151] The present invention is also directed to a process that practically eliminates requirements for compression or recompression of gas prior to returning a resulting product

-37-

Lean LNG in its liquid form after processing the Rich LNG. In this embodiment, the process comprises the steps comprise of:

[00152] (a) Pumping (22) Rich LNG (0-RICH LNG) from anywhere near to atmospheric pressure of for example about 10 PSIG pressure and about -260° F up to 700 PSIG or more (stream 1) depending on composition and conditions of the feed LNG to be processed.

[00153] (b) Heating the LNG (stream 1A) in an LNG exchanger 30 in a manner so that it is still maintained in a liquid condition (stream 2), in cross exchange with a warmer and methane rich vapor stream (8) received back from a processing column 50, particularly overhead vapor stream (7) from further downstream of the LNG exchanger 30.

[00154] The warmed Rich LNG (2, 3) is channeled to degasser equipment 40 ranging from a simple T Pipe device to a substantially liquid full vessel if desired.

[00155] (c) The degassed liquid (4) from the degasser 40 is flowed to a point in the upper section of the processing column 50.

[00156] (d) The processing column 50 operates at a pressure commensurate but selectable along with the full range of pressure of the pumped Rich LNG (1, 1A) from atmospheric to 700 PSIG, and anticipated above if the equilibrium conditions of the processing column 50 fluids allows separation.

[00157] (e) The processing column 50 produces a methane rich overhead vapor stream (7) from the top section of the column and an essentially NGL stream (6, 0-NGL) from the bottom section of the column 50.

[00158] (f) The processing column 50 receives when required in its operating mode a cold Lean LNG stream (10B) at a point in the column calculated for a particular combination of Pressures and fluid compositions to enhance its C2 Recovery or its C2 Rejection mode of operation.

[00159] (g) The Rich LNG feed (3,4) is similarly calculated for its best feed point in the column 50, starting at the top section of the column, with any associated gas (5) from the degasser 40 similarly calculated (though as demonstrated in the HYSYS result Tables included here under normal C2+ recovery mode no vaporization is anticipated (Tables 1,2,3) – except where a mode of operation such as in Ethane Rejection it may be incidental to the mode of operation (Tables 4, 6).

-38-

[00160] (h) The processing column 50 is demonstrated here working successfully with 10 theoretical trays.

[00161] (i) The column 50 has at least one contemplated reboiler 60 connected to the column for providing heat to the column operation.

[00162] (j) At least one C2+ enriched liquid stream is drawn from the column (6) and heated in the reboiler prior to returning the boiled stream (6C) back to the column 50.

[00163] (k) Optionally, a stream (6A) for recycle to the column 50 could be drawn from the reboiler 60 directly or the Product NGL Stream (6) and recycling that stream to the degasser 40 or at a point in the column 50.

[00164] (l) A stream (6A) for recycle to the column could be drawn from the NGL Product stream (6) and thence leaving the column arrangements as a NGL product stream (0-NGL).

[00165] (m) The drawn stream (6A) is pumped and recycled to the degasser 40 or directly at a point (not shown) calculated to fit the operation in the column 50 for a particular operational condition.

[00166] (n) The column 50 can be made to operate at various pressures (see streams 6 and 7) to meet the various interdependent parameters of the whole facility and desired performance.

[00167] (o) The overhead methane rich vapor stream (7, 8) from the column 50 is diverted to (8) the LNG exchanger 30 where it is condensed in cross exchange of heat with the cold Rich LNG feed (1A) where it ideally condenses up to 100% into Lean LNG liquid (9).

[00168] (p) The Lean LNG (9) is stored in a surge drum / receiving vessel 70, prior to pumping it (10A, 0-LEAN LNG) to storage or Pipeline at the required pressure.

[00169] (q) A part of the Lean LNG (0-LEAN LNG) product of (p) (e.g., stream 10D) as an option will be recycled to storage as part of a “roll over” control method.

[00170] (r) This is the contemplated introduction of a technique and method for control of storage LNG “Roll Overs”.

[00171] (s) From (q) it is contemplated spreading the part of the Lean LNG product (10D) as a further and part product of the Lean LNG flash of product of the Lean LNG

product of this invention which can be recycled to storage as part of a “roll over” control method.

[00172] (t) In the storage section 20 it is contemplated spreading a processed or cooled LNG product (10D) within or above a stored quantity of LNG as part of a “roll over” control method via jets or spargers (20C, 20B) relative to the LNG level 20A in the storage tank 20.

[00173] (u) Further it is contemplated sparging a processed product of this disclosed process, within (via jet(s)/sparger(s) 20B) or above (via jet(s)/sparger(s) 20C) a stored quantity of LNG having an LNG level 20A in the tank 20 as part of a “roll over” control method via jets or spargers.

[00174] (v) One option is also contemplated for sparging a processed LNG product comprising of vapor and/or liquid flash of this disclosed process, within or above a stored quantity of LNG as part of a “roll over” control method via jets or spargers.

[00175] The present invention provides advantages and features distinguishing over the systems of the prior art. For example, a flexible and streamlined NGL and C2+ extraction/rejection process is disclosed and as embodied for processing liquid, rich or virgin (feed composition) LNG (hereinafter called RLNG or also called VLNG) and essentially producing one, a liquid lean LNG (hereinafter may be called LLNG) product and second, liquid NGL (hereinafter also referred to as NGL) product(s) without the need of any of the typical compression or expansion work equipment, which however can be an optional part(s) of other embodiments (which as shown in one embodiment here, as an optional item (which further as demonstrated here requires 0 (zero) compression power, meaning there is no need for one in the instances shown with 0 (zero) horsepower). The present invention is a pressure flexible - source/(feed composition) flexible - product flexible - operation flexible - equipment and capital wise economic system/method/process.

[00176] In this process, a liquid phase hydrocarbon stream such as in this instance rich LNG (termed RLNG, or until its composition is affected/changed, called virgin LNG (VLNG)) which is rich in heavier/(less volatile) hydrocarbon components than such as in this instance methane, is introduced into the system as liquid phase VLNG from storage or transport system/pipeline. The liquid phase VLNG is introduced either in pressured state to or thence pumped (in a VLNG Pump) up to a pressure that as part of the inventive device/method/process supports the whole system's various parts' operating pressures as part of this inventive system. The stream is then passed via a heat exchanger (herein sometimes

-40-

called LNG exchanger, essentially a single exchanger but optionally more or a network of exchangers) while keeping it all as part of this inventive process essentially maintained in the heat exchanger(s) in a liquid phase with the controlled applied pressures (variable). The heat exchange takes place by picking up heat from the processing column (further downstream) vapor overhead stream (OVHD). The stream is then directed to a back pressure holding valve/device/(or back pressure from a downstream equipment/column) to maintain the stream in liquid phase. The stream is then sent to a mixer/separator/vessel/device (here called degasser) wherein the VLNG can be degassed of inert or light(er) components such as hydrogen, nitrogen, H₂S, CO₂, etc., but not limited to and thereafter, the gas stream is connected up to the processing column as gas/vapor inlet to the column. The gas stream can also be mixed with other optional streams.

[00177] Optionally (essentially depending on the mode of operation of the flexible system presented), a composition and enthalpy change can be effected to the VLNG via the degasser by optionally mixing with another stream which changes the state/composition of the feed VLNG prior to feeding (FEED) it to the processing/fractionation column. The stream is then directed to a column (wherein the vapor/liquid streams from the degasser can be connected at any number(s) of optimal feed locations on the fractionation column either as VLNG feed or changed composition feed stream or vapor and liquid streams. Pumping could be added where needed to pump the feed liquid to the processing column.

[00178] The present disclosure teaches pumping the RLNG from storage temperature and pressure to any particular pressure up to (and beyond if required) any process mixture thermodynamic critical pressure required in the column.

[00179] No stream division is contemplated in the present disclosure, as taught in the prior art, where of splitting of RLNG is contemplated for reflux purpose to column either from upstream of exchanger(s) or downstream of exchangers.

[00180] No vaporization of feed with original LNG composition is contemplated with the present process; the present teaching being to maintain the feed Rich LNG stream in its liquid form below its bubble point traversing the heat exchanger as an essentially sub-cooled liquid. No vaporization of Feed Rich LNG is "required" (vs. prior art) in any heat exchanger for feed or a portion of feed to the Column.

[00181] The use of a degasser is important in some modes of operation/flexibility.

-41-

[00182] Column pressures ranging from 50 psig to 1600 psig and more particularly 100 psig to 700 psig and even more particularly ranging from 400 psig to 700 psig with embodiments of close to 550 psig to 600 psig are included within the scope of the present disclosure.

[00183] No Rich LNG feed split is taken pre/post exchanger(s) (as noted in other prior art teachings) for purpose of feeding to column as column cold reflux and feed streams. Unlike with the prior art, no splits are taken of Feed Rich LNG stream to separately feed Column as feed and reflux.

[00184] In the present disclosure, the feed location on the column varies according to the mode of operation dictated from the hybrid functionality of the column described herein.

[00185] Various of the following columns can be emulated in various proportions effectively making this a hybrid column with various degrees of functions of the following column configurations with additional pressure swing functionality or pressure variability functionality to match most effective use of this process of disclosure: Distillation column; Extractive distillation column; Reboiled absorption column; Absorption column; Stripping column; Refluxed stripping column; and Reboiled stripping column.

[00186] The teachings of the present disclosure of employing a bottoms reboiler arrangement does not rule out the use of side exchangers and stream optimizations integrated to the column for heat distribution or recoveries; and further particularly heat recovery integration of bottoms discharge stream.

[00187] Column temperatures are managed by managing the bottoms exchanger temperature and feed stream properties to the column as demonstrated in the various streams to and from the column shown in TABLES 1A to 6A embodiments; stream 6 being indicative of bottoms temperature.

[00188] The present disclosure teaches obtaining almost 100% liquid lean LNG via condensing in the LNG exchanger against the cold of feed Rich LNG being processed as shown in the embodiments.

[00189] In one embodiment of the present process, the process can achieve up to 99% ethane recovery.

-42-

[00190] For NGL and C2-Recovery/Rejection modes of operation the present teachings can achieve industry/commercial Pipeline Specification ($<0.5\%$ C1 content, <600 psi TVP), NGL product without any further processing to achieve this result.

[00191] The present teaching can perform in “most” of the desirable modes liquid Lean LNG and high C2+ recoveries without any need for compression equipment.

[00192] A smaller Column is contemplated than other arts – the column of the present invention can require about 10 theoretical trays vs. others requiring about 20 trays.

[00193] The present invention teaches a versatile Column – a Column with hybrid-configurations and performance.

[00194] The present invention provide for economy of all equipment - number and duties.

[00195] One difference in the present process from the prior art is a reduction in the number of equipment and complexities of equipment/process/Degrees-of-Freedom. Another difference is that the present invention provides a system with ability to avoid temperature crosses in the LNG exchanger.

[00196] Yet another difference in this invention/process is that liquid state is maintained for the feed LNG (RLNG/VLNG) from storage to being warmed in exchanger(s) to liquid feed to degasser to column (optionally liquid feed can be connected directly to column) in contrast to the prior art teachings of vaporizing or partially vaporizing feed stream(s) or split parts of streams in exchanger/exchangers prior to feeding the VLNG stream(s) to the column. One manner in which to maintain the mixture in its liquid state is to maintain the mixture substantially or discernibly below bubble point. It is an object of the present invention to suppress as much as possible vaporization of rich LNG in the heat exchanger, and the present invention does not require the step of vaporizing at least a portion of the rich LNG prior to passing it to the fractionation column. Another difference in this inventive design/process is that liquid state is maintained for the feed LNG (RLNG/VLNG) from storage to being warmed in exchanger(s) to liquid feed to mixing-vessel/degasser wherein composition/enthalpy change is effected and the vapor/liquid outlets of mixing-vessel/degasser connected to optimal points in the processing column (optionally mixed composition effected feed stream can be connected directly to column without vapor/liquid separation, as an optional embodiment not fully shown here) in contrast to the prior art

-43-

teachings of vaporizing or partially vaporizing VLNG feed stream(s) or split parts of streams in exchanger/exchangers prior to feeding it to the column variously as non-composition-changed liquids/vaporized forms of VLNG.

[00197] Still another difference from the prior art is with one embodiment shown, a column bottoms liquid stream is recycled to the column via mixing in the degasser (vessel/device) changing the composition/enthalpy of the feed LNG which was up to this point VLNG (virgin LNG) prior to mixing with the warm bottoms stream. Yet another difference from the prior art is that the fractionation / processing column pressure is controlled to manage the LNG exchanger operation so that VLNG is kept in its liquid state while exchanging heat from the warmer processing column OVHD (overhead) stream. A further difference is that VLNG pump pressuring can be controlled flexibly depending on VLNG composition and state as a utility to make the overall system work as required in tandem with processing column pressure operation selection.

[00198] Another difference is that the column pressure is controlled/controllable to prevent “temperature crosses” in the LNG exchangers (“temperature crosses” being typical to other art) essentially making this inventive process as a whole practical/feasible and drastically reducing complexities of exchangers or network/banks of exchangers otherwise required to overcome “temperature crosses” or overcome undesirable/uneconomic exchanger design otherwise needed to overcome narrow “temperature approaches” or “temperature pinch” characteristics.

[00199] A further difference from the prior art is that the OVHD stream can be essentially fully condensed while exchanging heat and cooling down against the VLNG (or other optional streams/refrigeration not shown) without need for recompression prior to condensing as the LLNG liquid Product. Yet another difference is that the invention provides the ability for a purified lean LNG (LLNG) to be split and utilized as reflux in the processing column. The invention also provides, optionally, combinations of VLNG-Pump/Column/LNG-Exchanger/Bottoms-Recycle pressures/temperatures/compositions that can be adjusted to adjust slates of required product or BTU specifications for OVHD LLNG.

[00200] Another difference over the prior art is that the present invention provides, optionally, combinations of VLNG-Pump/Column/LNG-Exchanger/Bottoms-Recycle pressures/temperatures/compositions that can be flexibly adjusted (essentially utilizing VLNG-pump pressure, valves, processing column pressure with valves, stream recycles,

-44-

reflux) to adjust slates of required product specifications for NGL products for varying C2+ extraction/rejection content. Optionally, another difference and embodiment indicated is with certain combinations of operations vapor stream 11 from vessel 70 if generated which as gas stream can be condensed in a heat Exchanger CXG 90 after compression (80) and the condensed portion (12) can be mixed in with the LLNG (12D) or with the reflux (15A, 15B) to the column or degasser.

[00201] Yet another difference from the prior art is that optionally the stream 13 (or can be independent from process refrigeration/cooling stream 14A, 14B) can be used for condensing stream 12 from the compressor 80, and stream 13 may comprise another RLNG or other desirable stream to be processed or to enhance the operation that can be added in or mixed with the main VLNG or optionally connected directly to the column.

[00202] Any person skilled in the art or science, particularly one who is used to Process Engineering skills will, having had the benefit of the present disclosure, recognize many modifications and variations to the specific embodiment(s) disclosed. As such, the present disclosure, including examples, should not be used to limit or restrict the scope of the invention or their equivalents. Although embodiments have been shown illustrating operation of the processes of the present disclosure, those of ordinary skill in the art having the benefit of this disclosure could create other alternative embodiments that are within the scope of this invention. For example, with the benefit of the present disclosure, those of ordinary skill in the art will appreciate and understand that modifications and alternative embodiments to the process or method or system or improvements disclosed herein and comprise any feature described, either individually or in combination with any feature, in any configuration or individual steps or processes or combination of individual steps or processes for equipment design, operating, separating or recovering components of varying volatilities from Liquefied Natural Gas (LNG) or any other mix of hydrocarbons or other fluid mixes in a fluid phase.

[00203] References:

[00204] The following represents an exemplary list of U.S Patent references:

[00205] U.S. Patent No. 6,510,706 (Stone et al.) (January 28, 2003).

[00206] U.S. Patent No. 7,165,423 (Winningham) (January 23, 2007).

[00207] U.S. Patent No. 7,631,516 (Cuellar et al.) (December 15, 2009).

[00208] U.S. Patent No. 7,216,507 (Cuellar et al.) (May 15, 2007).

[00209] U.S. Patent No. 7,010,937 (Wilkinson et al.) (March 14, 2006).

[00210] U.S. Patent Publication No. 20080098770 (Ransbarger) (May 1, 2008).

[00211] U.S. Patent Publication No. 20090221864 (Mak) (September 3, 2009).

[00212] All references referred to herein are incorporated herein by reference as providing teachings known within the prior art. While the apparatus and methods of this invention have been described in terms of preferred embodiments, it will be apparent to those of skill in the art that variations may be applied to the process and system described herein without departing from the concept and scope of the invention. All such similar substitutes and modifications apparent to those skilled in the art are deemed to be within the scope and concept of the invention. Those skilled in the art will recognize that the method and apparatus of the present invention has many applications, and that the present invention is not limited to the representative examples disclosed herein. Moreover, the scope of the present invention covers conventionally known variations and modifications to the system components described herein, as would be known by those skilled in the art. While the apparatus and methods of this invention have been described in terms of preferred or illustrative embodiments, it will be apparent to those of skill in the art that variations may be applied to the process described herein without departing from the concept and scope of the invention. All such similar substitutes and modifications apparent to those skilled in the art are deemed to be within the scope and concept of the invention as it is set out in the following claims.

WE CLAIM:

1. A process for separating and recovering ethane and heavier hydrocarbons from LNG comprising the steps of:
 - a. providing an undivided feedstock stream containing Rich LNG wherein the Rich LNG is in liquid form from a storage tank or other source, the Rich LNG comprising C1 and C2+ hydrocarbons, the Rich LNG having an ambient storage temperature and pressure;
 - b. pressurizing the feedstock Rich LNG from storage pressure up to a desired pressure;
 - c. pumping the feedstock Rich LNG into the cool side of a heat exchanger, the heat exchanger having a cool side and a hot side;
 - d. heating the feedstock Rich LNG within the heat exchanger while maintaining the feedstock Rich LNG below its bubble point to avoid vaporization while in the heat exchanger;
 - e. directing the undivided feedstock Rich LNG feed stream from the heat exchanger to a processing column, the column comprising one or more stream entry ports along the height of the column to permit directing the stream into the column at one or more desired entry locations along the height of the column;
 - f. generating in the column a desired mixture comprising an overhead gas stream comprising lighter hydrocarbon products and a desired bottoms liquid stream comprising heavier hydrocarbon products;
 - g. directing the overhead gas stream from the column to the hot side of the heat exchanger;
 - h. cooling and condensing the overhead gas stream against the cold Rich LNG feedstock stream to form, in whole or in substantial part, a liquid comprising Lean LNG product stream, any remaining incidental uncondensed overhead gas stream remaining as a gas;

-47-

- i. directing the condensed product stream from the hot side of the heat exchanger to a receiving vessel;
 - j. pumping the liquid Lean LNG product from the receiving vessel to a desired location;
 - k. directing the bottoms liquid stream from the column to one or more reboiler arrangements;
 - l. heating the bottoms liquid stream in the reboiler;
 - m. returning at least a portion of the heated bottoms stream to the column, the column being further outfitted with one or more heated bottoms stream entry ports along the height of the column to permit directing the heated bottoms stream into the column at one or more desired heated bottoms stream product entry locations along the height of the column;
 - n. discharging the column bottoms stream directly from the column or from the reboiler and transferring the bottoms stream to a desired location; and
 - o. transferring any gas in the receiving vessel to a desired location.
2. The process of claim 1 wherein the desired pressure of step (b) is dictated by any downstream process steps involving the heat exchanger, and/or is dictated by critical pressure properties of the desired gas and liquid mixture in the column.
3. The process of claim 1 wherein the step of maintaining the feedstock Rich LNG below its bubble point to avoid vaporization while in the heat exchanger is achieved by regulating the pressure in the heat exchanger to maintain the Rich LNG in its liquid phase with no vaporization.
4. The process of claim 1 further comprising the steps of:
- a. directing the feedstock Rich LNG from the heat exchanger through a valve and into a degasser,
 - b. directing the liquid stream from the degasser into the processing column, the column being further outfitted with one or more degasser liquid stream entry ports along the height of the column to permit directing the degasser liquid

-48-

stream into the column at one or more desired degasser liquid product entry locations along the height of the column, and

- c. directing any gas stream in the degasser to the column, the column being further outfitted with one or more degasser gas stream entry ports along the height of the column to permit directing the degasser gas stream into the column at one or more desired degasser gas product entry locations along the height of the column.
5. The process of claim 4 wherein a portion of the column bottoms stream is directed to the degasser to warm the feedstock and alter the composition of the total feed to the column.
 6. The process of claim 4 comprising the additional steps of recovering heat from the column bottoms stream.
 7. The process of claim 1 wherein the NGL product comprises a desired high or low percentage of ethane.
 8. The process of claim 1 wherein the Lean LNG stream is directed to a storage facility or to further processing to vaporize the Lean LNG.
 9. The process of claim 1 wherein at least some of the Lean LNG stream is directed to the column, the column being further outfitted with one or more Lean LNG stream entry ports along the height of the column to permit directing the Lean LNG stream into the column at one or more desired Lean LNG product entry locations along the height of the column.
 10. The process of claim 1 comprising the additional steps of:
 - a. directing at least some of the Lean LNG stream into one or more additional heat exchangers,
 - b. heating the Lean LNG within the one or more heat exchangers while maintaining the Lean LNG below its bubble point to avoid vaporization while in the heat exchanger,

-49-

- c. directing the Lean LNG from the heat exchanger through a valve and into a degasser or other vessel,
 - d. directing the liquid stream from the degasser or other vessel to a desired location, and
 - e. directing any gas stream in the degasser or other vessel to a desired location.
11. The process of claim 10 comprising the additional steps of directing the liquid stream from the degasser or other vessel into another heat exchanger arranged in series relationship and repeating the steps of claim 10.
12. The process of claim 1 wherein the Lean LNG stream is directed to a Rich LNG feedstock storage containing a level of Rich LNG feedstock, the storage further comprising one or more jet or sparger systems located along the height of the storage to permit introduction of the Lean LNG stream into the storage either above and/or within the level of stored Rich LNG feedstock.
13. The process of claim 1 wherein the Lean LNG stream is directed to any stored source of LNG, wherein it is sparged into the stored source of LNG at a desired location.
14. The process of claim 1 wherein any gas phase in the receiving vessel is transferred to a compressor wherein the gas phase is compressed and then the compressed gas is directed to a desired location.
15. The process of claim 14 wherein the compressed gas is directed into a heat exchanger wherein the compressed gas is condensed to form a full or partial condensate Lean LNG, the condensate then being directed to a desired location.
16. The process of claim 15 wherein the condensate Lean LNG stream is directed to a storage facility.
17. The process of claim 15 wherein at least some of the condensate Lean LNG stream is directed to the column, and introduced into the column via the one or more Lean LNG stream entry ports to permit directing the Lean LNG stream into the column at one or more desired locations along the height of the column.

-50-

18. The process of claim 15 wherein the heat exchanger is cooled by an external refrigeration stream.
19. The process of claim 15 wherein the heat exchanger is cooled by a second LNG stream.
20. The process of claim 4 wherein a second cold LNG stream is introduced directly into the degasser to mix with the feedstock Rich LNG.
21. The process of claim 19 wherein the second cold LNG stream is introduced directly into the column, the column being further outfitted with one or more LNG stream entry ports along the height of the column to permit directing the LNG stream into the column at one or more desired LNG stream product entry locations along the height of the column.
22. The process of claim 1 wherein the step of cooling and condensing the overhead gas stream against the cold Rich LNG feedstock stream does not form any incidental gas.
23. The process of claim 1 wherein the discharged bottoms stream comprises up to 99% of the C2 hydrocarbons in the Rich LNG feedstock and substantially all of the C3+ as a NGL, the NGL product additionally meeting without any further processing close to or substantially a Pipeline Quality Specification of $\leq$ or $< 0.5\% \text{v C1}$.
24. The process of claim 1 wherein the discharged bottoms stream comprises an NGL Product of substantially with TVP of up to $< 400 \text{ psig}$, up to C1 $\leq$ or $< 0.5\% \text{v}$, up to 51%mol C2 or more fraction.
25. The process of claim 1 wherein the Rich LNG feedstock comprises between 1% mole C2 to that exceeding 40 to 50 mole % C2.
26. The process of claim 1 wherein the process runs in a high "ethane recovery" (90% or more) mode to recover up to 99% ethane and substantially 100% propane.
27. The process of claim 1 wherein the column comprises about 10 theoretical trays.
28. The process of claim 1 wherein substantially no tail gas (gas from condensed overhead stream) is formed even where there is as low as to 1% C2 in the feedstock.

-51-

29. The process of claim 1 wherein NGL of Pipeline Quality specs is produced, even when the system is operating in deep high ethane extraction (90% plus) mode.
30. The process of claim 1 wherein the column is configured and integrated in a multitude of operability and functional configurations selected from the group consisting of distillation columns, extractive distillation columns, reboiled absorption columns, absorption columns, lean oil absorber columns, fractionation columns, stripping columns, refluxed stripping columns, and reboiled stripping columns.